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COAL GASIFICATION STUDY.(U)

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CIVIL ENGINEERING LABORATORY
Naval Construction Battalion Center
Port Hueneme, California

Sponsored by
NAVAL FACILTIES ENGINEERING COMMAND

COAL GASIFICATION STUDY

April 1977

An Investigation Conducted by

BECHTEL CORPORATION
San Francisco, California

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The general problem of providing fuel gas for Navy base facilities is studied. The intent is: first, to provide designs of a coal gasification plant producing 6×10^9 Btu/day reactor output, based on two types of reactors; second, to conduct parametric studies leading to means for the costing of similar plants operating on different feedstocks; and third, to provide

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a method for estimating the change in boiler rating which must follow the substitution of fuel gas for either oil or coal firing.

The performance and economics given are based on conceptual design methods. The economic results allow comparison of fuel-gas and fuel-oil costs on the basis of the Navy's method of analyzing costs using "Economic Analysis Handbook," NAVFAC P-442, 1975. The costs are the sum of all future outlays discounted to the present but allowing escalation at different rates for utilities and feedstock over a 25-year production period.

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SUMMARY

The general problem of providing fuel gas for Navy base facilities is studied. The intent is:

- To provide designs of a coal gasification plant producing 6×10^9 Btu/day reactor output, based on two types of reactors.

- To conduct parametric studies leading to means for the costing of similar plants operating on different feedstocks.

- To provide a method for estimating the change in boiler rating which must follow the substitution of fuel gas for either oil or coal firing.

- To present the latter methods in a handbook.

The objectives are met in this report and associated handbook.*

The performance and economics given are based on conceptual design methods. The economic results allow comparison of fuel-gas and fuel-oil costs on the basis of the Navy's method of analyzing costs. The costs are the sum of all future outlays discounted to the present but allowing escalation at different rates for utilities and feedstock over a 25-year production period.

Fuel gas made from $25/ton bituminous coal costs about $1.60 per 10^6 Btu, while fuel oil costs about $1.86 per 10^6 Btu, according to the Navy's method.

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*Civil Engineering Laboratory. Contract Report CR 77.014, "Coal Gasification Study Handbook," Bechtel Corporation, San Francisco, California, 1977.

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Section 1

INTRODUCTION

This study examines a coal-derived fuel, specifically industrial gas, as one of the alternate fuels available to a Navy base.

The national energy program calls for a combination of conservation and development in our efforts to meet the energy needs. The dependence on foreign oil has grown to a critical level despite the potential for instability in both price and supply. While conservation can be counted on to relieve a portion of a lapse in supply, the final need is to find alternatives to foreign oil as a source of energy.

Coal is a candidate resource of the required order of magnitude, offering perhaps 100 years of energy supply for the expected consumption rates and satisfying an equally important condition - reliability. If the arrival of a substantial coal/energy industry accomplishes nothing more than competition with foreign oil, it will encourage more reasonable and dependable character to the supply of foreign oil.

The gasification of coal has been the subject of development for decades, beginning with the production of the first town gas at the turn of the century. About 50 years ago the producer reactor became available in an alternate form, the fluidized bed reactor. About 40 years ago, the so-called entrained solids reactor was developed.

More recently, as a result of the shortage of oil, acceleration in development in coal conversion to gas took place. The oil embargo

indicated the danger of a dependence on foreign source for the U.S. supply. Also, the concern of environmentalists has brought an increased stringency to the cleaning of gases and to the prevention of their forms of pollution.

NAVY BASE FUEL SUPPLY

The gaseous fuel considered in this report is used primarily to fire boilers and to serve auxiliary heaters, not necessarily to serve as fuel for domestic housing. The fuel is considered as a replacement for natural gas or oil and must meet pollution regulations. Reasons for consideration of fuel gas as an alternative to direct-fired coal are the ease with which it can be distributed and fired and the versatility of its applications. The fuel supply operation is simplified.

At the same time, a number of problems must be faced that are usually the concern of the utilities producing the conventional fuels and which arise in departing from their fuels.

These problems are:

- Maintenance of a reliable supply of fuel.
- Conversion of the fuel consuming facilities to use of a new fuel.
- Adjustment of the ratings of these facilities, where needed.
- Capacity for the turndown of fuel production.
- Operation of a logistical system to support operation of the fuel plant.
- Provision of adequate removal of sulfur particulates from the fuel gas.
- Disposal of liquid and solid wastes.

It must further be assumed that these operations are performed with a cost that is competitive with utility or other supplier charges.

FUEL COSTS

The production of fuel at low cost depends on operation at optimal conditions, taking advantage of the economies of large-scale production.

Development work has long been concerned with the economy of gasifying coal, and the latest projects (second generation) promise economies that have been estimated at perhaps 25 per cent over first generation operation. This advantage, however, must compete with the continuing escalation in the cost of new plants.

The cost of gaseous fuel, however, will be acceptable in almost any case if a serious shortage of conventional fuel develops. The value of coal will then escalate. It is important to find the least expensive coal-derived fuels for the long term *in this event*.

ENVIRONMENTAL FACTORS

Environmental effects of process plant operation are now responsible for as much as one-third of the capital cost of the plant and the cost of its operation. The regulation of emissions may become gradually tighter, and it is prudent to consider more stringent regulations than now exist.

A variety of processes is now available for removing sulfur from gases. The choice of an appropriate process depends on the pressure level at which the gas is to be delivered, the amount of CO_2 allowable in the product fuel gas, and the concentration level to which sulfur is to be reduced.

The deposition of wastes has become another technical problem area. The processing of coal inevitably generates large quantities of solid

waste, which must be sent to permanent sealed dumps. Drainage from the dump area must be monitored, collected, and processed before being discharged to a water body. The dump must be adequately covered to restore a portion if not all of the original character of the land surface.

ALTERNATIVES

A number of alternative fuels would be worthy of some consideration as competitors of fuel gas. The factors that bear indirectly on the comparison are:

- Reliability of supply
- Compatibility with user equipment
- Ability to meet environmental regulations

Liquified petroleum gas is one such candidate. It is available at perhaps \$0.50 per 10^6 Btu premium over the price of natural gas (available at \$2.20/10^6 Btu). It is available in substantial quantities and requires little capital investment in plant. It has the disadvantage of dependence on foreign sources.

Another coal-derived fuel is cleaned coal, which has been treated physically or chemically to remove mineral-bound sulfur. The present technology indicates that about one-half of the total sulfur (mineral plus organic) can be removed from a large portion of presently available coal. The treatment would bring a large portion of present coal production within the present limits imposed by federal regulations. Ultimately, this portion would be depleted, but in all likelihood would be augmented by the additional coal reserves made eligible by improved methods of treatment. The indicated cost of sulfur removal by physical or chemical methods now considered moderate-term technology is of the order of \$0.50/per 10^6 Btu.

The direct combustion of any available coal followed by stack gas cleanup will soon be a well demonstrated route to clean energy under the present regulations. The cost of cleanup is likely to be in the vicinity of $0.50-0.75 per 10^6 Btu.

GLOSSARY

1.	Acid Gas	— Hydrogen sulfide and carbon dioxide, alone or in mixture with other components of minor concentration.
2.	Combustibles	— Considered in this report to be the gaseous components of useful fuel value, exclusive of H_2S.
3.	Fire Tube Boiler	— A boiler in which the hot gases pass through the main exchanger tube bundle "in-tube."
4.	Fuel Gas	— Considered in this report to be the gases from the reactor containing H_2S, but exclusive of water vapor.
5.	Hot Fuel Gas	— Considered in this report to be all the hot raw gases from the reactor. Cool fuel gas is considered to be the same, but leaving the waste heat recovery section at perhaps 350°F.
6.	High Heating Value, HHV	— Considered in this report to be the higher heating value of the 100 lb coal used as the basis for analysis.
7.	Lean Solution	— Gas scrub solution which is relatively spent in absorption capacity; i.e., solution which is high in concentration of absorbed component.
8.	Product Gas	— The gas exported from the plant for consumption.
9.	Quench Water	— The low pressure steam injected for quench purposes in the entrained solids reactor.
10.	Savings Investment Ratio, SIR	— The ratio of the present value of savings offered by the subject investment program divided by the present value of the investment required for instituting that program.
11.	Service Factor	— The percentage of time for which a plant, or section of a plant, is operating usefully.
12.	Sour Gas	— A gas containing hydrogen sulfide and other minor sulfur compounds.

13. Sweet Gas — A gas free of sulfur components.

14. Tail Gas — The residual gas emerging from the incinerator following the Claus plant; containing small amounts of SO_2 to be scrubbed out before release to the atmosphere.

15. Waste Heat Recovery — Generally, the recovery of sensible heat from hot gases but more specifically in the present analysis considered as the recovery of heat from the raw hot reactor gases.

16. Water Tube Boiler — A boiler in which the main exchanger carries water "in-tube."

Section 2

STATEMENT OF THE PROBLEM

The Navy wishes to exploit the developing technology of coal gasification in helping implement the national energy program. Coal gasification could provide fuel for use on the Naval bases. The Navy desires specifically to find the cost of fuel gas, typically available from present or near future technology. Furthermore, it desires a method for evaluating costs for other cases in which the characteristics of the coal and gas and/or emission regulations are different from those of the nominal case. Finally, it wishes to have a method for adjusting the derating of boilers to reflect conversion to fuel gas. The plant design is to represent the best tradeoffs in cost consonant with the available technology in gasification and gas cleanup. The contractor is to rely on his own resources primarily. The specific tasks are given below. The depth of analysis is to be conceptual in nature.

GASIFICATION STUDIES

Task 1 — Nominal Plant Design

The Navy desires the design for a specific plant for gasification of a typical coal, involving the best set of operating conditions for the production of cheap gas. Also, the Navy desires the capital and operating costs of this nominal case. The plant should be designed for two of the three types of gasifers presently regarded as commercially available, operating in both the oxygen blown and air blown modes. The candidate types of gasifiers are:

- Fixed bed
- Fluidized bed
- Entrained solids

Two of these types are to be selected for analysis, with the important categories of capital and operating costs to be delineated.

Task 2 — Parametric Analysis

A major objective of the project is to provide the Navy with the means for finding costs for other situations with different constraints. Thus, the treatment of the following parameters is to be accounted for:

- Characteristics of coal
  - Type
  - Heating value
  - Moisture content
  - Ash content
  - Sulfur content
- Pressure of the gas treatment section
- Heating value of gas
- Emission limit on sulfur in gas

The performance of the plant and capital costs of the modules are to be specified, as are also the separate contributions to operating costs. Finally, the cost of gas is to be derived. This procedure is to be presented in a handbook, whose use will permit the evaluation of different coals and various sulfur emission levels.

BOILER DERATING (TASK 3)

The objective is to show how to estimate the approximate effects on existing Navy shore-based facility boiler performance when changed from present fuels to coal-derived gas fuels. A method is to be provided for the types of equipment:

- Watertube boilers (capacity to 200,00 lb/hr steam)
- Firetube boilers (capacity to 10,000 lb/hr steam)
- Coal-fired boilers (stoker feed)

The contributions to heat transfer from connection and radiation are to be analyzed and the results provided in the form of curves or equations.

Section 3

TECHNICAL APPROACH

The sequence of study is parallel with the sequence of tasks given in the preceding section.

The initial step is to identify the gasifier types and modes of operation to be studied. Next, the performances and costs of typical plant systems are calculated. These "nominal" performances provide a basis from which behavior at modified conditions can be found by extrapolation. This effort comprises Task I.

The study of parametric effects is undertaken as Task II. The large volume of computation is managed with a computer program. Certain coefficients are derived for use in constructing a "short form" analysis. The latter procedure supports hand adjustment of nominal system performance to reflect arbitrary changes in parameters.

A second hand procedure is developed for treating the boiler derating problem. This work is Task III.

Both hand procedures are presented in a handbook separate from the final report.

GASIFIER TYPES

Three types of gasifiers are considered for the proposed Navy application:

- Fixed-bed reactor
- Entrained-solids reactor
- Fluidized-bed reactor

Two of these are chosen in a preliminary study to be analyzed in depth in subsequent work. The choice is discussed in some detail in Appendix G, and the study was limited to the entrained solids and fluid bed types of reactors. Both types are operable in oxygen blown mode, but only one type is efficient in air blown modes of operation. Thus, the study is confined to three basic cases.

COMPUTER PROGRAM

The performance of a typical coal gasification plant is calculated on a module-by-module basis. The validity of analysis is verified by vendor information on the performance of equipment. The means for performing the analysis is the GASPLANT computer program, which defines the performances of the gasifier, the gas treatment section, the compressors, and the various auxiliaries that make up the rest of the plant. The program also provides for determining the cost of the installation, the operations, and the product gas. A Bechtel-sponsored project is providing this program.

NOMINAL PLANT, PARAMETRIC STUDIES

The analyses are then performed for many sets of conditions, but reflect only one type of coal. These sets represent the successive variations in:

- Type of gasifier and mode of blast
- Pressure at which raw fuel gas is treated
- Sulfur content of feed coal
- Sulfur content of exit gas

Each set represents one computer run.

For each gasifier/blast-mode system, a number of sets is studied to derive response of the cost of the process system to changes in gas pressure and sulfur levels.

The additional variables noted in Section 2, i.e., coal characteristics, are not treated to find general effects in the computer runs. They are satisfactorily included as parameters, however, in a short form analysis noted below.

One pressure condition is finally chosen as the economic optimum for the plant system for each gasifier-type and mode of operation (air or oxygen blast). Together with the preselected values for the sulfur contents noted above, these conditions define three nominal plants as shown in Table 3-1. Case 1 performance is fully reported; Cases 1 and 2, partially.

Table 3-1

DEFINITION OF NOMINAL PLANTS

	Nominal Plant		
	Case 1	Case 2	Case 3
Blast Mode	O_2/steam	O_2/steam	Air/steam
Gasifier Type(1)	Entrained solids	Fluidized solids	Fluidized solids
Sulfur Content in Coal (%)	2	2	2
Sulfur Emission in Gases (lb SO_2 per 10^6 Btu HHV of coal)	1.2	1.2	1.2
Gas Treatment Pressure (single value for all cases)	Optimum	Optimum	Optimum

Note:

(1) The fixed bed gasifier was ruled out of consideration for reasons of complexity in gas cleanup. Also, the entrained solids gasifier is operated only with oxygen blast for reasons of efficiency; the air must be preheated if used.

The remaining analyses (computer runs) represent off-design conditions serving the function of providing correction coefficients. Such coefficients support the adjustment of performance and costs in the "hand procedure" described below. These runs are termed "variants" below.

These analyses are presented below in the form of cost data and curves.

HAND ANALYSIS

The interrelations and effects of the listed parameters on costs in the problem are to be evaluated in the short procedure to be presented in the handbook. While correlation of the gas cleanup parameters is handled on an empirical basis, the treatment of the remaining ones is on a more fundamental, thermodynamic basis. Thus, heating value, coal type, and moisture and ash content are assessed in a primarily stoichiometric study of the partial combustion of the coal. This requires the ultimate analysis and ash softening temperature. The adiabatic reaction temperature of the gasification reaction is above or below the ash softening temperature, reflecting the slagging or non-slagging character of the reactor operation.

It should be pointed out that the parameters listed above are not necessarily independent of each other. It is desirable to find the heat effects generated in the gasification process and the thermochemical energy remaining in the product fuel gas. They connect the parameters in an unambiguous way. The analysis leads to the approximate temperature of the reactor, the amount of gas produced, and the amount of steam made in waste heat recovery. This information is about all that is needed to describe gasification performance. The analysis forms the first portion of the hand procedure, and leads to costs.

COSTING (HAND METHOD)

The costing of the plant modules is handled by relating the capacity or size of the module to that of the corresponding unit in the nominal plant design and making an appropriate extrapolation of cost from that of the nominal unit. This procedure should be valid, since the extrapolation will usually be small.

The evaluation of unit product cost is performed on the basis of the discounted cash flow method. It observes the Navy position toward depreciation, which is to write off the full cost of the plant immediately as a one-time contribution to the cost of gas.

SUMMARY

In summary, the project effort is two-fold in its treatment of the plant performance.

- A series of computer runs is made to find the performance and the costs of the plant as it generates a fuel gas from a given coal. The behavior of the gas treating section and the resultant effects on gas costs are studied. A simplified correlation between the parameters applying to this section and the resultant costs is developed, and a nominal design is provided for each gasifier/blast mode-type.
- A simplified thermodynamic analysis of the gasifier section is developed. To this is added the correlation above, accounting for the gas treatment section. Together these analyses relate plant operation parameters to the overall capital and operating costs on a simplified basis, comprising a hand method. The hand method is compared with a computer run in terms of results.

The analyses are conceptual in depth of detail.

BOILER DERATING

Three types of boilers are included in this study:

- Water tube boilers up to 200,000 lb/hr of steam production
- Fire tube boilers up to 10,000 lb/hr of steam production
- Stoker feed, coal-fired water tube boilers up to 200,000 lb/hr of steam production

Since no specific designs for these three boiler types were specified, it is necessary to define a boiler configuration that is representative of each of the three boiler types. The representative (or standard) boiler configuration for each boiler type is assumed to be rated at the maximum capacity stated above. It is assumed that the rating change factors defined for the maximum capacity boiler for each boiler type will generally apply to lower capacity boilers of that type.

Additional general assumptions for these boiler designs can be made. Saturation pressures at 150 psig correspond to steam temperatures of 366^{o}F. These conditions along with the age of boilers support assumptions of natural steam circulation in the boilers and that air pre-heaters are not used with the boilers.

Boiler derating factors for the boilers are obtained in this study by evaluating the "first order" effects of fuel change to coal-derived gas. The effects on boiler efficiency, steam capacity, and combustion product flow rates are evaluated for the fuel change variations of:

- Combustion reaction temperature
- Emissivity factor
- Gas flow rate

Details of the analytical method used are included in the appendix to this report.

Section 4

PROCESS DESCRIPTION

GENERAL

The gasification plant, Case 1, is designed to produce 6 billion Btu/day of raw fuel gas by gasifying entrained coal with oxygen and steam. The fuel gas produced is a medium Btu gas with an HHV of 300 Btu/scf and is to be used, after being treated, as a substitute for natural gas. Case 1 is described fully below; Cases 2 and 3 are presented only in the form of the process stream tabulations, Appendix D.

FACILITY, CASE 1

The gasification plant consists of the following sections:

- Coal Handling and Preparation
- Air Separation (Oxygen Plant)
- Gasification
- Fuel Gas Compression and Expansion
- Fuel Gas Treating
- Claus Plant and Tail Gas Treating

These are shown in the process block flow diagram, Figure 4-1.

GASIFICATION

The gasification plant is designed to gasify 350 tons per day of dry pulverized coal to produce 6 billion Btu/day of medium Btu fuel gas. The unit consists of the following sections:

- Gasification
- Waste Heat Recovery
- Dust Removal

Gasification

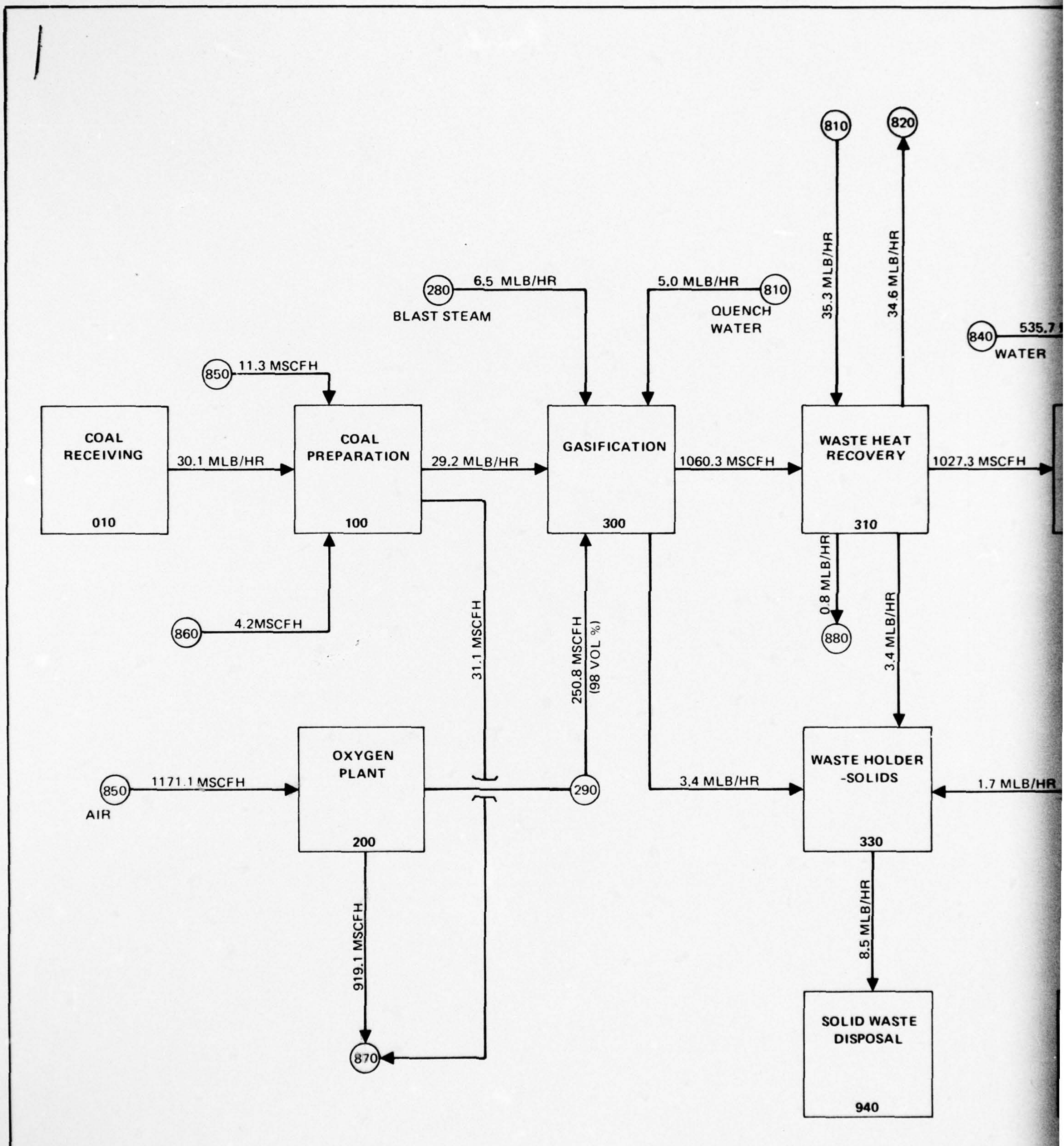
Dry, pulverized coal is fed to a mixing head by means of a screw feeder. At the mixing head, a mixture of steam and oxygen entrains the pulverized coal. Entering the gasifier, carbon in the entrained oxygen-steam stream is exothermally oxidized, producing a high temperature flame zone in the region of 3500°F. Endothermic reactions between the carbon and steam substantially reduce the flame temperature while continuing in the process to oxidize carbon and produce additional hydrogen. Also the reactants are quenched with water spray for added temperature control.

The gasifier is an entrained solids reactor and is shaped like two spheroidal cones centrally welded together at the bases. It has a double-walled shell, and the annulus between the inner and outer shell is water-cooled and connected to a steam separation drum. The low-pressure steam generated in the double-walled shell is used as the process steam which enters the gasifier through the mixing heads. Pressure at the reactor is at 35 psia; hot raw fuel gas leaving the reactor is at 2350°F.

Ash in the fuel is liquefied in the high temperature flame zone of the gasifier. Forty percent of the ash falls down the gasifier walls as molten slag and drains into a slag quench tank. The remainder of the ash leaves the gasifier as fine fly ash entrained in the exit gas.

Waste Heat Recovery

Hot gases at 2350°F emerging from the gasifier are sent through a waste heat boiler to recover the sensible heat in the hot fuel gas. Cool gas leaves the waste heat boiler at 366°F. High-pressure steam,


COAL RECEIVING
010
30.1 MLB/HR
COAL PREPARATION
100
850
11.3 MSCFH
860
4.2MSCFH
29.2 MLB/HR
31.1 MSCFH
280
6.5 MLB/HR
BLAST STEAM
810
5.0 MLB/HR
QUENCH WATER
GASIFICATION
300
1060.3 MSCFH
250.8 MSCFH
(98 VOL %)
290
810
35.3 MLB/HR
820
34.6 MLB/HR
WASTE HEAT RECOVERY
310
1027.3 MSCFH
0.8 MLB/HR
880
3.4 MLB/HR
840
535.7
WATER
850
AIR
1171.1 MSCFH
OXYGEN PLANT
200
919.1 MSCFH
870
3.4 MLB/HR
WASTE HOLDER -SOLIDS
330
1.7 MLB/HR
8.5 MLB/HR
SOLID WASTE DISPOSAL
940

2

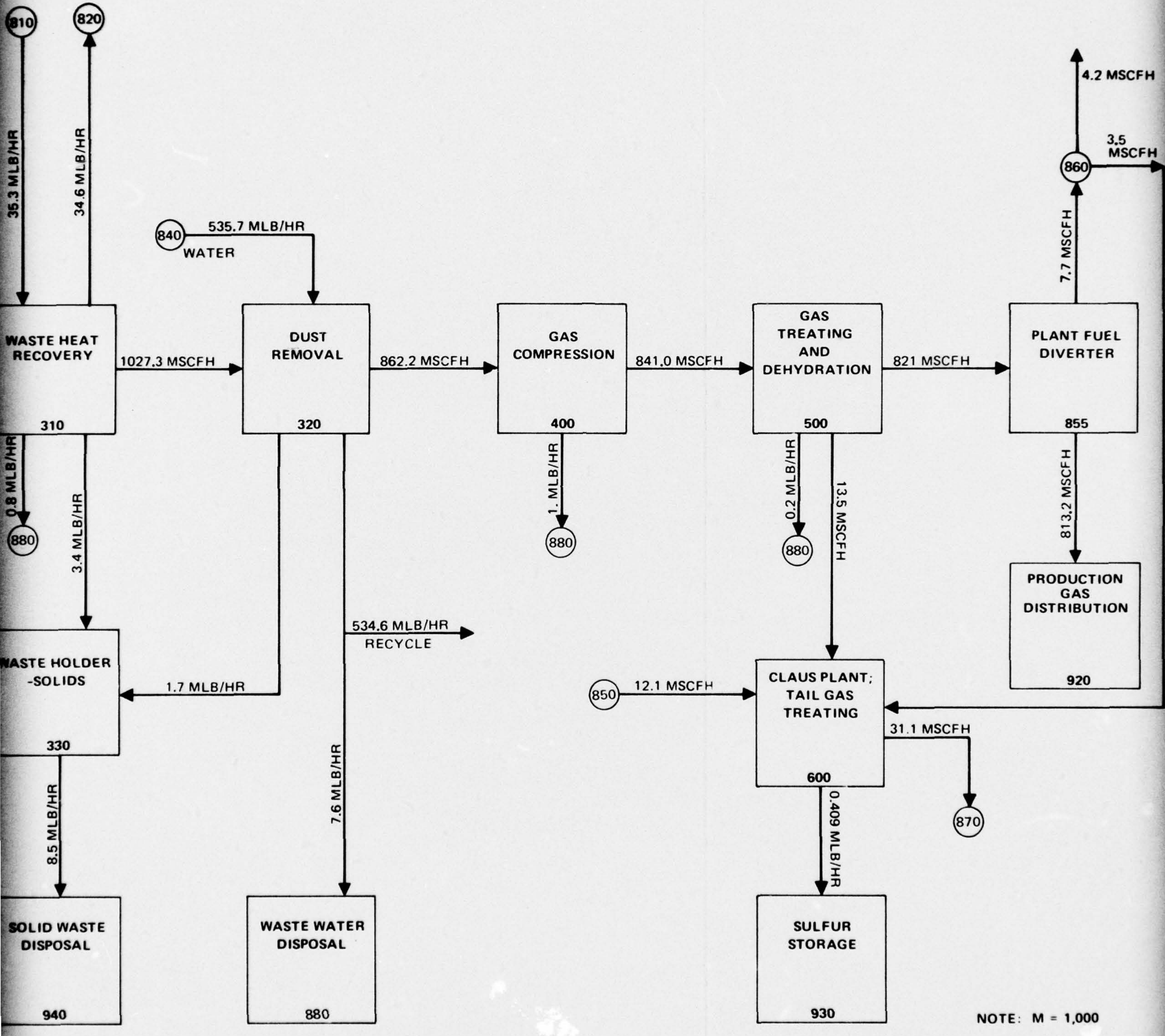


Figure 4-1. BLOCK FLOW DIAGRAM, CAS

900°F, 1,055 psia, is generated at the rate of 34,600 lb/hr. Part of this steam is letdown in pressure and is used as steam blast, in gas treating, and in pyrol vapor recovery. High pressure steam which is in excess is used for air compression.

The fuel gas exiting the waste heat boiler is washed and cooled in a spray washer.

About 70 percent of the heavy particulate matter is removed, and the fuel gas is cooled down to 104°F.

Dust Removal

Subsequent cleaning of the cool fuel gas is accomplished in two Theissen disintegrators. The remaining particulate matter in the fuel gas is removed; clean fuel gas leaves the second disintegrator. Carry-over moisture in the fuel gas is then removed by means of a mist eliminator. The fuel gas is sent to an electrostatic precipitator prior to compression.

FUEL GAS COMPRESSION, EXPANSION

Compression

The particle-free fuel gas coming out of the dust removal section is compressed from 35 psia to 165 psia. This is accomplished in a two-stage centrifugal compressor. The compressor is equipped with an intercooler, a postcooler, and knockout vessels. An electric driver is used to drive the compressor. This gas passes to the gas treatment section.

Fuel Gas Reheat and Power Recovery System

The system is designed to recover power in the expansion of fuel gas exiting the gas treating section (described below) from 150 psia to 40 psia. Reheating of the fuel gas to support expansion is accomplished by heat exchange with:

- Compressed warm gas prior to treating
- Raw gas in the waste heat recovery section

Three pieces of equipment are used in the process:

- Sweet Fuel Gas/Compressed Warm Gas Exchanger. This is a shell-and-tube exchanger which allows partial heating up of the 100°F sweet gas exiting the gas treating section by means of heat exchange with the hot gas discharging from the second stage compression section.

- Sweet Fuel Gas/Raw Gas Exchanger. This is a finned tube exchanger located at the bottom of waste heat recovery section. It heats up the partially heated sweet gas flowing through the the tubes to the temperature required for expansion. Hot raw gas in the waste heat recovery section serves as the heating medium and flows down through the fins.

- Expander. Recovers power by expanding the reheated (nominal) 150 psia sweet fuel gas to 40 psia, 100°F.

A low temperature source (compressed gas, 280°F) has to be utilized to partially heat up the 100°F sweet fuel gas. This is done because hot raw gas in the waste heat recovery section could not be solely utilized to bring about the heating of the sweet gas since it contains a great amount of water vapor. Appreciable cooling down of the raw gas would result in condensation, which must be avoided.

For further details, please see Figure 4-2.

GAS TREATMENT

The raw, dust-free fuel gas contains a substantial concentration of H_2S and a small amount of COS, the sum of which is usually enough to require extraction of sulphur if the environmental regulations are to be met. It is then necessary to provide treatment, which may include

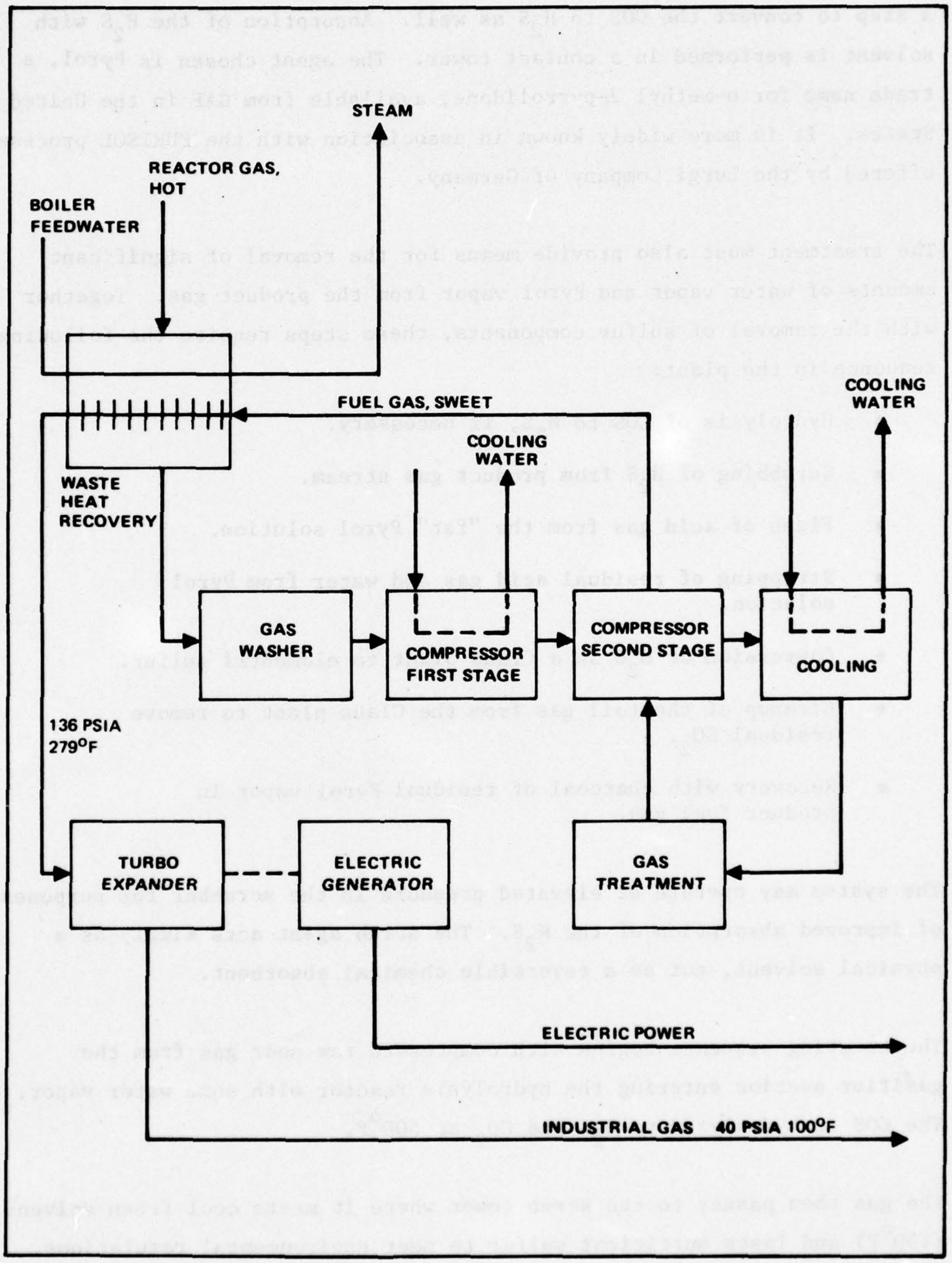


Figure 4-2. POWER RECOVERY BY EXPANSION OF FUEL GAS, CASE 1

a step to convert the COS to H_2S as well. Absorption of the H_2S with solvent is performed in a contact tower. The agent chosen is Pyrol, a trade name for n-methyl 2-pyrrolidone, available from GAF in the United States. It is more widely known in association with the PURISOL process offered by the Lurgi Company of Germany.

The treatment must also provide means for the removal of significant amounts of water vapor and Pyrol vapor from the product gas. Together with the removal of sulfur components, these steps require the following sequence in the plant:

- Hydrolysis of COS to H_2S, if necessary.
- Scrubbing of H_2S from product gas stream.
- Flash of acid gas from the "fat" Pyrol solution.
- Stripping of residual acid gas and water from Pyrol solution.
- Conversion of H_2S in a Claus plant to elemental sulfur.
- Cleanup of the tail gas from the Claus plant to remove residual SO_2.
- Recovery with charcoal of residual Pyrol vapor in product fuel gas.

The system may operate at elevated pressure in the scrubber for purposes of improved absorption of the H_2S. The scrub agent acts simply as a physical solvent, not as a reversible chemical absorbent.

The treating sequence begins with compressed raw sour gas from the gasifier section entering the hydrolysis reactor with some water vapor. The COS is hydrolyzed to H_2S and CO_2 at 500^oF.

The gas then passes to the scrub tower where it meets cool fresh solvent (100^oF) and loses sufficient sulfur to meet environmental regulations. At the same time, the gas is dehydrated enough to fall in dew point to

around $0^{o}F$. Further, the gas is treated in a charcoal bed to absorb a low content of Pyrol vapor (∿ 1.5 mm), recovering perhaps 95 percent. The charcoal is loaded with pyrol vapor to some level under 10 percent by weight before it is bypassed. The absorption cell is regenerated by heating with a recirculating stream of hot fuel gas. The cycle for each of the three cells used in rotation is at least one-half day.

The product gas now passes to the turbo expander for energy recovery.

The H_2S absorbed by the solvent in the scrub tower is accompanied by a certain amount of CO_2. The solution passes to a flash chamber where the greater part of the gases desorb. The remainder is removed in a stripping tower, and all the acid gas then passes through a water condensation sequence before approaching the Claus plant. Concurrent water rejection from the stripped solvent takes place in a fractionator, providing the dryness necessary to support the dehydrating action of pyrol on fuel gas at the front end of the system, i.e., the scrub tower.

COAL HANDLING AND PREPARATION

The coal handling system consists of unloading, crushing, and storage operations. Coal is delivered to the plant site by means of railroad cars. A small car unloader equipped with a transfer device and an 18-inch wide belt conveyor unloads and transfers the coal into a storage bin from where the coal is fed into a crusher. The crusher selected is Hammermill type rotary crusher capable of reducing the 6" x 0" coal to 99 percent passing 1/4 inch size. Crushed coal is transferred into a silo for storage by means of a variable speed, 18-inch wide belt conveyor at the rate of 25 tons per hour. The silo is 22 feet in diameter, 44 feet in height, and has a capacity of about 15,500 cubic feet of active coal.

From the silo, the coal is fed to a bowl mill, where it is dried and pulverized simultaneously. The drying medium, either hot flue gas or

fuel gas combusted with excess air, is circulated through the mill. Coal is dried from 5 percent to 2 percent moisture content and is pulverized to about 70 percent through 200 mesh. The dried pulverized coal is then sent through a cyclone separator to a storage bunker. From the storage bunker, it is fed to a service bin, then to a feed bin, and finally to a screw feeder. From the screw feeder, it is fed to the gasifier at the rate of 350 tons per day.

OXYGEN/AIR PLANT

The air separation plant receives 1,171,000 scfh of air for oxygen generation. This is accomplished by compression, liquefaction, and fractionation of the air feed to produce 251,000 scfh of O_2 (98 vol % O_2).

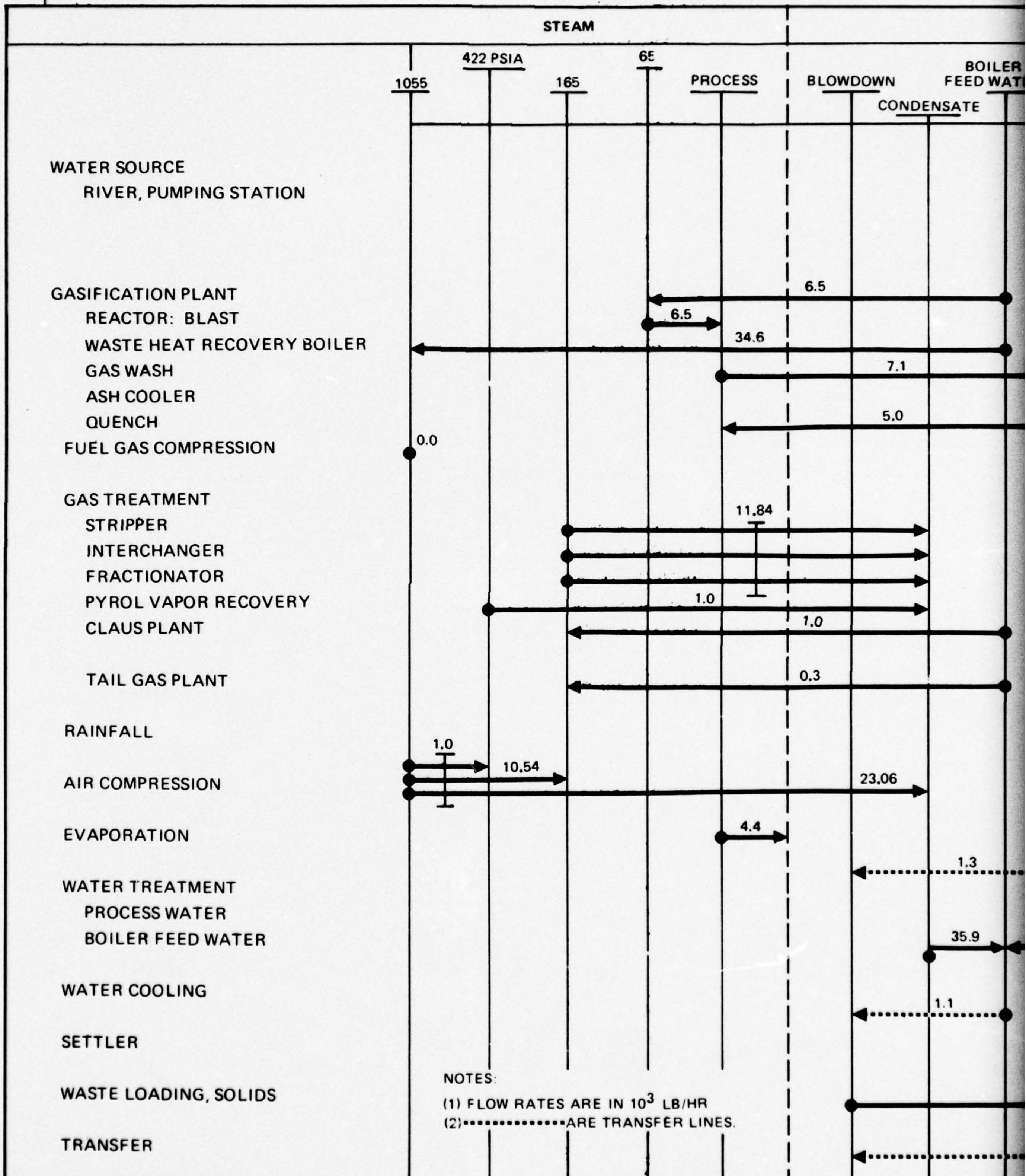
Ambient air is compressed to 165 psia using a centrifugal compressor. The compressor is equipped with intercoolers, a postcooler, and liquid knockout vessels. A steam turbine utilizing excess high pressure steam from the waste heat recovery section is used to partially drive the compressors. This steam turbine is supplemented by an electric driver to provide the remaining power.

Liquefaction and fractionation of the air is accomplished in the "cold box group." Product oxygen is then sent to an oxygen compressor where the oxygen is compressed to the required gasification pressure.

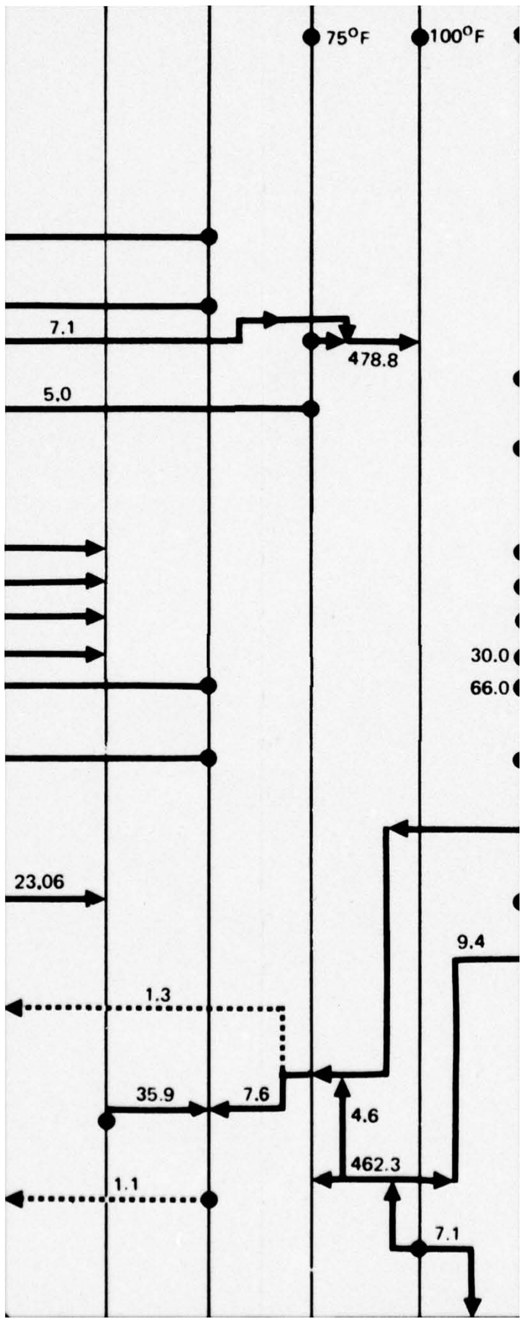
STEAM/WATER CIRCUIT, UTILITIES

The steam/water circuit shows the distribution of steam, process water, and cooling water in the gasification plant. It is composed of the following manifolds (see Figure 4-3):

- Steam
- Condensate/BFW



STEAM
422 PSIA
65
1055
165
PROCESS
BLOWDOWN
BOILER
FEED WAT
CONDENSATE
WATER SOURCE
RIVER, PUMPING STATION
GASIFICATION PLANT
REACTOR: BLAST
WASTE HEAT RECOVERY BOILER
GAS WASH
ASH COOLER
QUENCH
FUEL GAS COMPRESSION
GAS TREATMENT
STRIPPER
INTERCHANGER
FRACTIONATOR
PYROL VAPOR RECOVERY
CLAUS PLANT
TAIL GAS PLANT
RAINFALL
AIR COMPRESSION
EVAPORATION
WATER TREATMENT
PROCESS WATER
BOILER FEED WATER
WATER COOLING
SETTLER
WASTE LOADING, SOLIDS
TRANSFER
6.5
6.5
34.6
7.1
5.0
0.0
11.84
1.0
1.0
0.3
1.0
10.54
23.06
4.4
1.3
35.9
1.1
NOTES:
(1) FLOW RATES ARE IN 10³ LB/HR
(2) ARE TRANSFER LINES.


75°F
100°F
7.1
478.8
5.0
30.0
66.0
23.06
9.4
1.3
35.9
7.6
4.6
462.3
1.1
7.1

- Process water
- Cooling water
- Raw water supply

Steam

High pressure steam, 1,055 psia, is generated in the waste heat recovery section of the gasifier and is sent to a distribution header. This steam generation supplies the various steam demands within the system at different pressure levels, namely:

- 65 psia steam — blast to gasifier
- 165 psia steam — gas treating
- 422 psia steam — pyrol vapor recovery

High pressure steam, which is in excess after meeting these steam requirements, is utilized for air compression.

Condensate/BFW

All of the steam generated, except that which is used as process steam (65 # steam blast) is sent to condensate. This condensate undergoes treatment and is then sent to the BFW header. Makeup BFW from rainwater supply and/or river water undergoes the same kind of treatment.

Process Cooling Water

Cooling water for condensing the water vapor in the fuel gas and for cooling down the fuel gas from 350°F to 104°F is fed in at 75°F and leaves at 100°F.

Cooling Water

The different sections of the plant that require cooling water are;

- Gasifier section (ash cooler)
- Fuel gas compression
- Air compression

Cooling water enters at $75^{o}F$ and leaves at $100^{o}F$.

Raw Water Supply

There are two sources of raw water:

- Rainfall
- River

Rainwater, which is stored in tanks, is the primary source of any makeup water requirements. River water is used as supplement.

Water Treatment

Makeup water in the form of rainfall and river water has to undergo treating prior to its being used either as process water or boiler feed water. Process cooling water return has to pass through a settler to remove dust particles present in it. This is done before the cooling water is recycled.

WASTE DISPOSAL

The solid wastes from the gasifier are delivered by belt conveyor from a collection point at the gasifier module to a separate holding bin. Solids are then discharged from the bin periodically to the box of a dump truck, hauled to a port facility nearby, and loaded on a barge.

The solids will be damp, after dust-control water spraying, and, in one case, as a result of the ash solids being drain-wet after quenching at the gasifier module. The solids are dumped at sea where their fundamentally basic pH condition and soluble mineral content will contribute to the nutrient content of the sea water.

The liquid wastes are from water treatment operations and are discharged through a small pipeline to the nearby sea or otherwise to a natural stream. The wastes will contain such mineral solids as carry over from a settling basin and also salt from ion exchange regenerators. Underflow sludge from the settler will be sent to the dry ash dampening operation.

PLANT LAYOUT

The future site of the gasification plant is a 26 acre land tract, as shown in Figure 4-4. It accommodates a plant in which process flow is clockwise, beginning with arrival of coal.

Coal is delivered to the plant site by means of railroad cars. It is unloaded and transferred to a coal storage (stockpile) which holds 10,600 tons of coal. This is done by the use of car unloaders equipped with a transfer device and an 18-inch-wide belt conveyor. From the coal storage, another belt conveyor feeds the coal to a crusher. Crushed coal is conveyed to a storage bin at the rate of 25 tons per hour. From the storage bin, the coal is fed to a ball mill where it is dried and pulverized simultaneously. Dried, pulverized coal is sent to a bin. It is then charged to the gasification plant which consists of a gasifier, a waste heat recovery section, and a dust removal section. Particle-free cool gas exiting the dust removal section passes to the compressor and then to the scrubber to remove the H_2S and CO_2 present. The treated fuel gas then undergoes expansion at the turbo-expander to recover some power. It is sent to pipeway and off to market.

The acid gas (H_2S and CO_2) from the scrubber continues to the Claus plant where the H_2S is burned to SO_2. Tail gas from the Claus plant goes to a tail gas treating plant. Treated tail gas is sent to pipeway, then to stack; sulfur, a product of acid gas treating, is sent to a sulfur storage, and thence back to the railway for shipping.

ENERGY DISTRIBUTION

The energy distribution chart shows (Figure 4-5) the flow of energy in the gasification plant. Its purpose is to provide a complete analysis of the "cold gas" efficiency, the "gasifier" efficiency, as well as some other efficiencies described in the latter part of this report. It also shows the amount of power derived from excess WHB steam, if any, and the electric power needed to supplent it for the overall plant requirement.

Five sections of the gasification plant are exhibited in the figure; energy redistribution in three sections is described below.

- Gasification

  Energy source

  - HHV of coal
  - Sensible/latent heat of air/oxygen/steam blast

  Energy redistribution

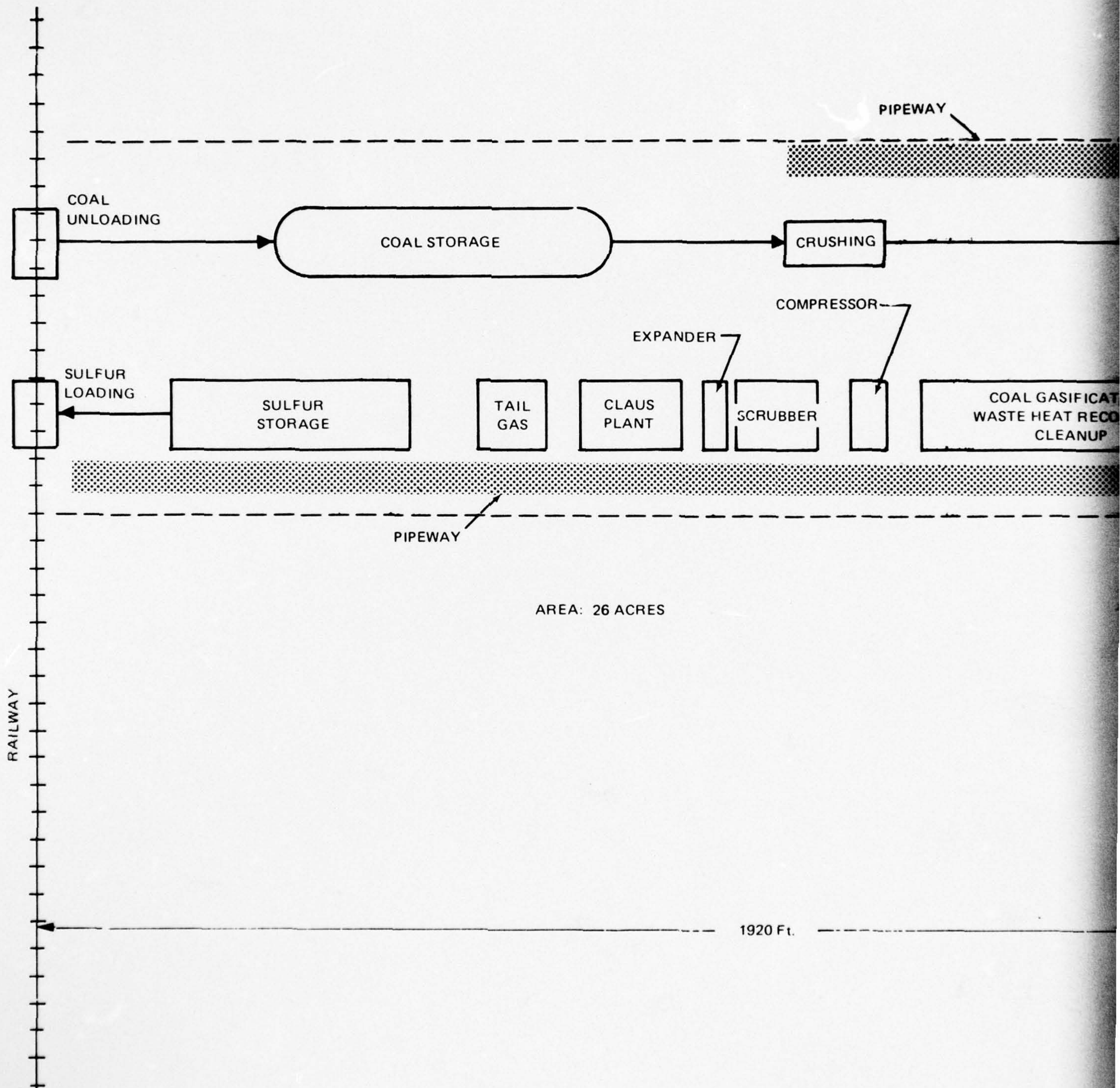
  - Calorific heat of raw fuel gas
  - Sensible/latent heat of hot raw gas; slag

- Fuel gas compression and treating

  Energy source

  - Calorific heat of particle-free cool gas

1


PIPEWAY
COAL
UNLOADING
COAL STORAGE
CRUSHING
COMPRESSOR
EXPANDER
SULFUR
LOADING
SULFUR
STORAGE
TAIL
GAS
CLAUS
PLANT
SCRUBBER
COAL GASIFICAT
WASTE HEAT RECO
CLEANUP
PIPEWAY
AREA: 26 ACRES
RAILWAY
1920 Ft.

2

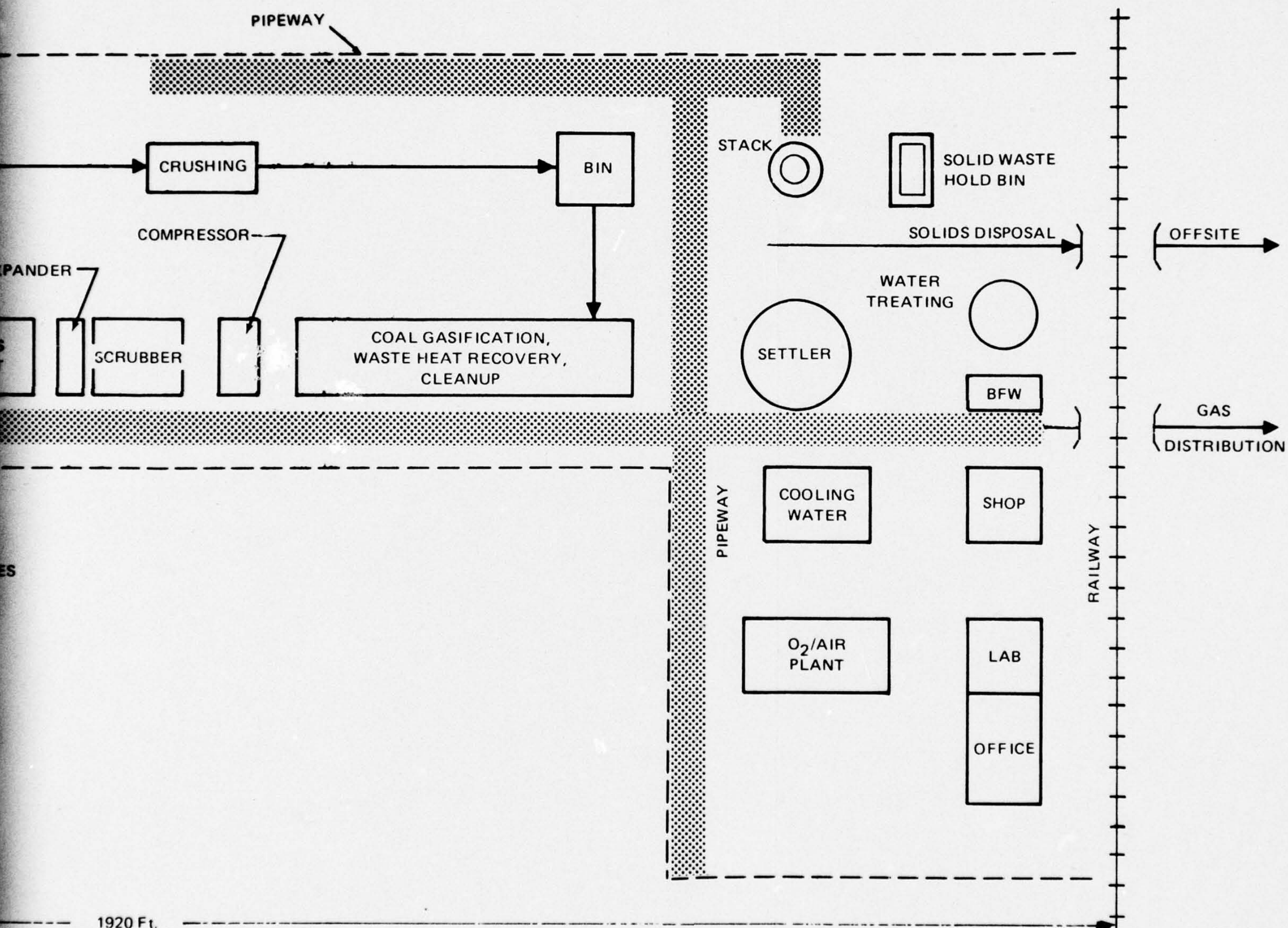
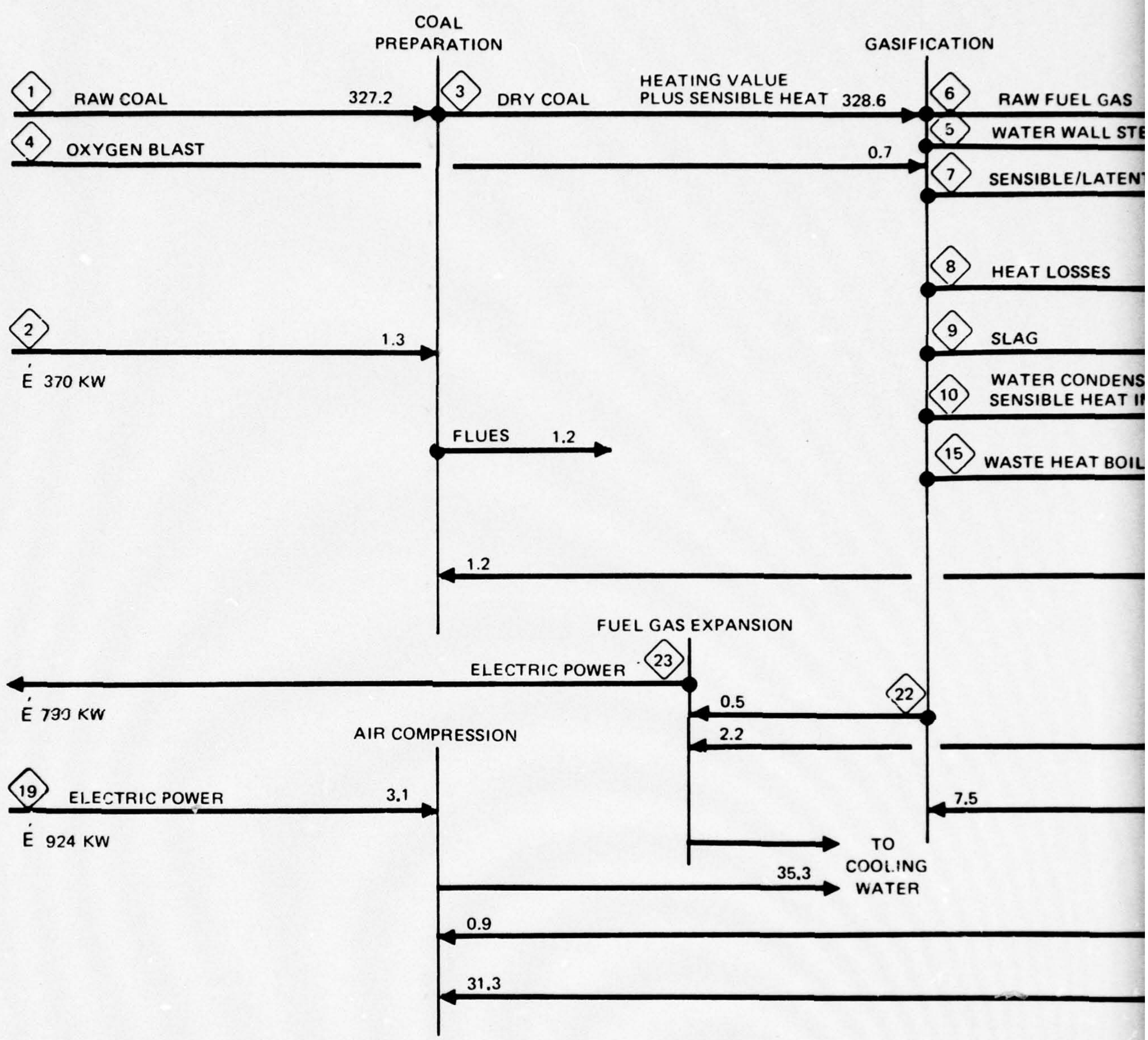


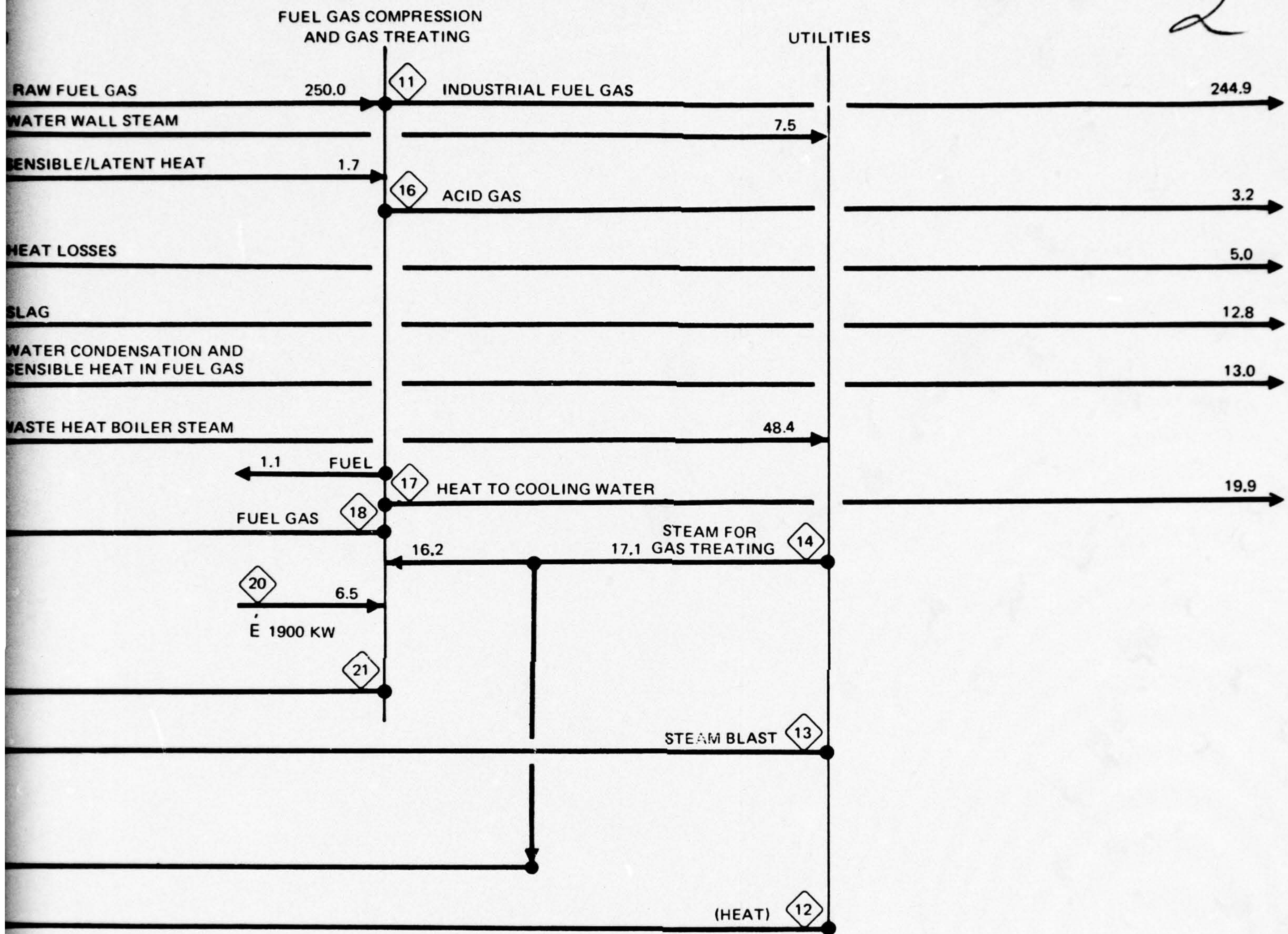
Figure 4-4. GASIFICATION PLANT LAYOUT



NOTE: VALUES ARE IN 10^6BTU/HR, EXCEPT AS OTHERWISE STATED. É = ENERGY RATE, KW

STREAM NUMBER	1	2	3	4	5
DESCRIPTION	RAW COAL CALORIFIC HEAT	ELECTRICITY	DRY COAL CALORIFIC SENSIBLE HEAT	AIR/OXYGEN SENSIBLE HEAT	WATER WALL STEAM
STREAM NUMBER	13	14	15	16	17
DESCRIPTION	STEAM BLAST	STEAM FOR GAS TREATING	STEAM GENERATION IN WASTE HEAT BOILER	ACID GAS HEAT	HEAT TO COOLING WATER AND CONDENSATION

2



5	6	7	8	9	10	11	12
ATER ALL TEAM	FUEL GAS CALORIFIC HEAT	SENSIBLE LATENT HEAT	HEAT LOSSES	SLAG HEAT	WATER CONDENSATION AND HEAT TO COOLING WATER	INDUSTRIAL FUEL GAS	STEAM FOR AIR COMPRESSION
17	18	19	20	21	22	23	
EAT TO OOLING ATER AND NDENSATION	FUEL GAS TO COAL PREPARATION	ELECTRICAL POWER TO AIR COMPRESSION	ELECTRICAL POWER TO FUEL GAS COMPRESSION	HEAT FROM FUEL GAS COMPRESSION	HEAT FROM HOT REACTOR GASES	ELECTRICAL POWER	

Figure 4-5. ENERGY DISTRIBUTION, CASE 1

- Sensible/latent heat of particle-free cool gas
- Enthalpy of steam to treating
- Heat of compression in fuel gas

Energy redistribution

- HHV of fuel gas to coal drying
- Heat to cooling water

• Air Compression

Energy source

- Steam power
- Electrical power

Energy redistribution

- Heat to cooling water

Efficiencies

• Ratio No. 1 (cold gas efficiency)

$$\frac{\text{HHV of product fuel gas}}{\text{HHV of coal}} = 75.0\%$$

• Ratio No. 2 (gasifier efficiency)

$$\frac{\text{HHV of raw fuel gas + steam generation}}{\text{HHV of coal + calorific/sensible heat input to gasifier}} = 90.8\%$$

The distribution of the total energy input in the system, i.e., the HHV of coal and electrical energy, to the major consumers within the plant is described below in terms of three ratios:

• Ratio No. 3 (gas treatment requirements)

$$\frac{\text{Q (Gas treatment + compression)}}{\text{Q (HHV of coal + } \sum \text{ Electric Energy)}} = 6.73\%$$

- Ratio No. 4 (energy to fuel gas)

$$\frac{\text{HHV of fuel gas}}{\text{HHV of coal} + \sum \text{Electric Energy}} = 72.7\%$$

- Ratio No. 5 (air compression)

$$\frac{\text{Energy for air compression}}{\text{HHV of coal} + \sum \text{Electric Energy}} = 10.5\%$$

PERFORMANCE

The performance of the CASE 1 nominal plant, developed from GASPLANT computer program runs is presented in several figures. These are the process block flow diagram, Figure 4-1; the steam water circuitry, Figure 4-3; and the energy distribution diagram, Figure 4-5. Further detail on the compositions of the process streams is given in Table A, Appendix D.

The performances of the remaining nominal plants, CASES 2 and 3, are given only in tabulations of stream flow rates and compositions in Tables B and C in Appendix D. The diagrams are essentially of the same structure for all plant cases. Equipment for each nominal plant case is listed in Appendix E.

SERVICE FACTOR

The plant operates 90 percent of each year.

OPERATING LABOR

Table 4-1 shows the distribution of manpower among plant functions.

Table 4-1

OPERATING LABOR

Operating Labor	Men/Shift
Solids Handling	1/2
Coal handling	
Solid waste disposal	
Sulfur disposal	
Gasification	1-1/2
Oxygen and Compression	1
Oxygen (or air) plant	
Compression plant	
Desulfurization, Deyhydration	
H_2S scrubbing and stripping, COS hydrolysis	1-1/2
Claus and tailgas-treatment	
Sorbent vapor recovery	
Utilities and Off-Plots	
Water clarification, BFW treatment	1-1/2
Cooling tower	
TOTAL	6

Chapter 5

RESULTS

GASIFICATION PLANT ECONOMICS

Nominal Plants

Investment costs, annual operating costs, and product gas unit costs are presented below for the three nominal plants.

Investment Costs. The investment costs are shown in Table 5-1. The tabulation includes construction and plant equipment costs for each of nine plant subsections, distributable costs associated with site operation during construction, and engineering costs and fee. The table also includes other nonrecurring costs, which cover process royalties, operating personnel training, spare parts inventory, initial charge of catalysts and chemicals, plant startup costs, and owner management and reporting costs. Each capital cost entry in Table 5-1 includes a contingency allowance for small additional cost items that cannot be identified without a complete detailed design. This contingency allowance is one-sixth of each cost item in the table. The contingency represents five percent of the gas unit cost.

The investment costs do not include the price of land or the cost of modifying or rearranging any existing equipment on any Navy site. Also, no allowance has been made for one-time costs for working capital, future plant salvage value, and changes in existing Navy site assets.

Annual Operating Costs. The annual operating costs for the three plants are shown in Table 5-2. The amount of coal and electricity needed is

Table 5-1

PLANT INVESTMENT COSTS FOR CASES 1, 2, AND 3
(Thousands of dollars)

Plant Section	Case 1	Case 2	Case 3
Coal Preparation	2,850	2,228	2,228
Oxygen Supply	7,064	5,992	1,774
Gasification	5,426	6,016	7,216
Compression, Gas Expansion	1,706	1,758	3,350
Desulfurization and Dehydration	889	904	1,622
Sulfur Recovery	1,332	2,701	1,780
Interconnecting Piping	1,825	1,353	1,492
Utilities	478	565	486
Waste Disposal	424	424	424
Direct Field Cost	21,994	21,941	20,372
Distributable Field Cost	2,195	1,352	1,628
Total Field Cost	24,189	23,293	21,000
Engineering, Home Office, and Fee	2,661	2,562	2,420
Total Construction Costs	26,850	25,855	24,420
Startup Costs	2,950	2,844	2,686
Total Capital Costs	29,804	28,699	27,106
Basis			
Blast Mode	O_2/steam	O_2/steam	Air/steam
Gasifier Type	Entrained solids	Fluidized solids	Fluidized solids
Sulfur Content in Coal, %	2	2	2
Sulfur Emission in Gases, e (SO_2) lb SO_2 per 10^6 Btu HHV of coal	1.2	1.2	1.2
Gas Treatment Pressure, optimum, psia	150	150	150

Table 5-2

ANNUAL OPERATING COSTS FOR CASES 1, 2, AND 3
(Thousands of dollars)

Cost Element	Case 1	Case 2	Case 3
Coal @ $25/ton	2,969	2,889	3,313
Purchased Electric Power @ $.030/kw-hr	673	736	245
Catalysts and Chemicals	80	80	80
Equipment, Supplies, Utilities	80	80	80
Operating Personnel	528	528	528
Maintenance Materials and Labor	800	800	800
Total Annual Operating Costs	5,130	5,113	5,046
Basis			
Blast Mode	O_2/steam	O_2 steam	Air/steam
Gasifier Type	Entrained solids	Fluidized solids	Fluidized solids
Sulfur content in coal, %	2	2	2
Sulfur Emission in Gases lb SO_2 per 10^6 Btu HHV of Coal	1.2	1.2	1.2
Gas Treatment Pressure, Optimum, psia	150	150	150

slightly different for each of the three plants. The other cost items are identical for the three plants.

<u>Product Gas Unit Costs</u>. Tables 5-3, 5-4, and 5-5 present the discounted average future cost of gas for the three gasification plants and compare the values with discounted future price of fuel oil. They also present a savings/investment ratio for each plant compared to the fuel oil alternative.

The product gas unit cost for each plant is the ratio of the plant project present value divided by the total Btu's of product gas produced by the plant. The resulting gas cost is expressed in $/million Btu. Because the cost calculated in this way is a heavily discounted future cost, it should not be compared with gas costs for current operations that are usual in commercial feasibility studies.

The present value for each plant project is the discounted sum of all cost cash flows over the life of the project. This sum includes initial investment costs and the recurring annual operating costs. The zero of time for the calculations is January 1977. The construction period lasts for three years beginning in January 1978 and ending in December 1980. The plants are assumed to produce gas continuously for 25 years from January 1, 1981 through December 3, 2005. Half the investment costs fall before the start of the third construction year (January 1, 1980). One-sixth, one-third, and one-half of the investment costs are assumed to fall in the first, second, and third years of construction, respectively.

In calculating the present values, discount factors were taken from the <u>Economic Analysis Handbook.</u>* The investment costs were multiplied by a one-time discount factor. Operating costs were multiplied by series sum discount factors that apply for continuous costs from the fifth to

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*<u>Economic Analysis Handbook</u>, P-442, Naval Facilities Engineering Command, 1975.

Table 5-3

PRODUCT UNIT COST AND SIR FOR CASE 1*

Line Number	Cost Element	Differential Inflation Rate	Project Year	Amount, Thousands of Dollars		Discount Factor	Discounted Cost, Thousands of Dollars
				One Time	Recurring		
(1)	First-Year Construction	+0	2	4 967		0.867	4 306
(2)	Second-Year Construction	+0	3	9 935		0.788	7 829
(3)	Third-Year Construction	+0	4	14 902		0.717	10 685
(4) •	Total Investment			29 804			• 22 820
(5)	Coal	+5	5-29		2 969	12.268	36 424
(6)	Electricity	+6	5-29		673	14.057	9 463
(7)	Operating Labor and Materials	+0	5-29		1 488	6.505	9 679
(8) •	Total Operating Costs				5 130		• 55 566
(9) •	Total Project Costs						• 78 386
(10)	Fuel Oil Alternative	+8	5-29		4 820	18.631	89 800
(11)	Energy Available over 25 years, billions of Btu						48 199
(12) •	Product Gas Unit Cost, $/million Btu (line 9 divided by line 11)						• 1.63
(13) •	Fuel Oil Alternative Unit Cost, $ million Btu (line 10 divided by line 11)						• 1.86
(14)	Savings/Investment Ratio, SIR = (line 10 - line 8)/line 4						1.50

Table 5-4

PRODUCT UNIT COST AND SIR FOR CASE 2*

Line Number	Cost Element	Differential Inflation Rate	Project Year	Amount, Thousands of Dollars		Discount Factor	Discounted Cost, Thousands of Dollars
				One Time	Recurring		
(1)	First-Year Construction	+0	2	4 783		0.867	4 147
(2)	Second-Year Construction	+0	3	9 566		0.788	7 538
(3)	Third-Year Construction	+0	4	14 350		0.717	10 289
(4) •	Total Investment			28 699			• 21 974
(5)	Coal	+5	5-29		2 889	12.268	35 442
(6)	Electricity	+6	5-29		736	14.057	10 354
(7)	Operating Labor and Materials	+0	5-29		1 488	6.505	9 679
(8) •	Total Operating Costs				5 113		• 55 475
(9) •	Total Project Costs						• 77 458
(10)	Fuel Oil Alternative	+8	5-29		4 825	18.631	89 896
(11)	Energy Available over 25 years, billions of Btu						48 184
(12) •	Product Gas Unit Cost, $/million Btu (line 9 divided by line 11)						• 1.61
(13) •	Fuel Oil Alternative Unit Cost, $ million Btu (line 10 divided by line 11)						• 1.86
(14)	Savings/Investment Ratio, SIR = (line 10 - line 8)/line 4						1.57

*See Table 5-7 for case identification.

Table 5-5

PRODUCT UNIT COST AND SIR FOR CASE 3

Line Number	Cost Element	Differential Inflation Rate	Project Year	Amount, Thousands of Dollars: One Time	Amount, Thousands of Dollars: Recurring	Discount Factor	Discounted Cost, Thousands of Dollars
(1)	First-Year Construction	+0	2	4 518		0.867	3 917
(2)	Second-Year Construction	+0	3	9 035		0.788	7 120
(3)	Third-Year Construction	+0	4	13 553		0.717	9 718
(4) ●	Total Investment			27 106			●20 755
(5)	Coal	+5	5-29		3 313	12.268	40 644
(6)	Electricity	+6	5-29		189	14.057	3 440
(7)	Operating Labor and Materials	+0	5-29		1 488	6.505	9 679
(8) ●	Total Operating Costs				4 990		●53 763
(9) ●	Total Project Costs						●74 518
(10)	Fuel Oil Alternative	+8	5-29		4 820	18.631	89 771
(11)	Energy Available over 25 years, billions of Btu						48 184
(12) ●	Product Gas Unit Cost, \$/million Btu (line 9 divided by line 11)						● 1.55
(13) ●	Fuel Oil Alternative Unit Cost, \$ million Btu (line 10 divided by line 11)						● 1.86
(14)	Savings/Investment Ratio, SIR = (line 10 - line 8)/line 4						1.73

CASE IDENTIFICATION

Basis	Case 1	Case 2	Case 3
Blast Mode	O_2/Steam	O_2/Steam	Air/Steam
Gasifier Type	Entrained Solids	Fluidized Solids	Fluidized Solids
Sulfur Content in Coal, %	2	2	2
Sulfur Emission in Gases, Lb SO_2 per 10^6 Btu HHV of Coal	1.2	1.2	1.2
Gas Treatment Pressure, Optimum, psia	150	150	150

twenty-ninth year of the project. Each such sum discount factor is the difference between the cumulative uniform series factor for years 1 to 29 and that for years 1 to 4 of the project. The factors shown in the tables for recurring costs are these differences.

Energy costs are assumed to escalate faster than ordinary general inflation over the next 30 years. The differential inflation rates used for Navy studies should be as follows (2):

- Coal +5%
- Electricity +6%
- Labor and Materials +0%
- Fuel Oil +8%

The discount rate of 10 percent is assumed in Tables 5-3, 5-4, and 5-5 as indicated on page 17 of Reference*. It is interesting to note that discount rate of 10 percent plus differential inflation rate of 8 percent is mathematically equivalent to a discount rate of 2 percent.

Low sulfur fuel oil would be an alternative to gas. All three gasification plant projects produce thermal values at substantially lower cost than that available from fuel oil burned over the same 25 years beginning in 1981. An appropriate January 1977 price for purchased fuel oil is \$2.50/million Btu.**

A second useful way to compare the plants with the alternative of burning fuel oil is to calculate the savings/investment ratio (SIR). This recognizes that burning fuel oil will require no initial investment. On the

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*"Energy Escalation Rates for Short Term Costing and Life-Cycle Costing," Naval facilities Engineering Command, 1976.

**"Assessment of Availability and Price of Fossil Fuels for Utility Purposes Through 1985," Hoff-Muntner Corporation, June 1975, page, 21.

other hand, the gasification plants would require initial investment. The SIR is the present value of the annual savings afforded by the gasification plants, divided by the present value of the investment for the plants. The present value of the annual savings is the difference between the present value of oil purchased over the 25 years and the present value of gasification plant operating costs for 25 years. When the SIR is greater than one, the savings will more than pay for the investment. The SIR turns out to be substantially greater than one for each of the three gasification alternatives. Accordingly, the construction of gasification plants is economically justified.

Parametric Sensitivity Analysis

The three nominal plants described in Section 4 were modeled using the GASPLANT computer program developed by Bechtel. Then six or seven off-nominal parametric variants were run and compared for each nominal case.

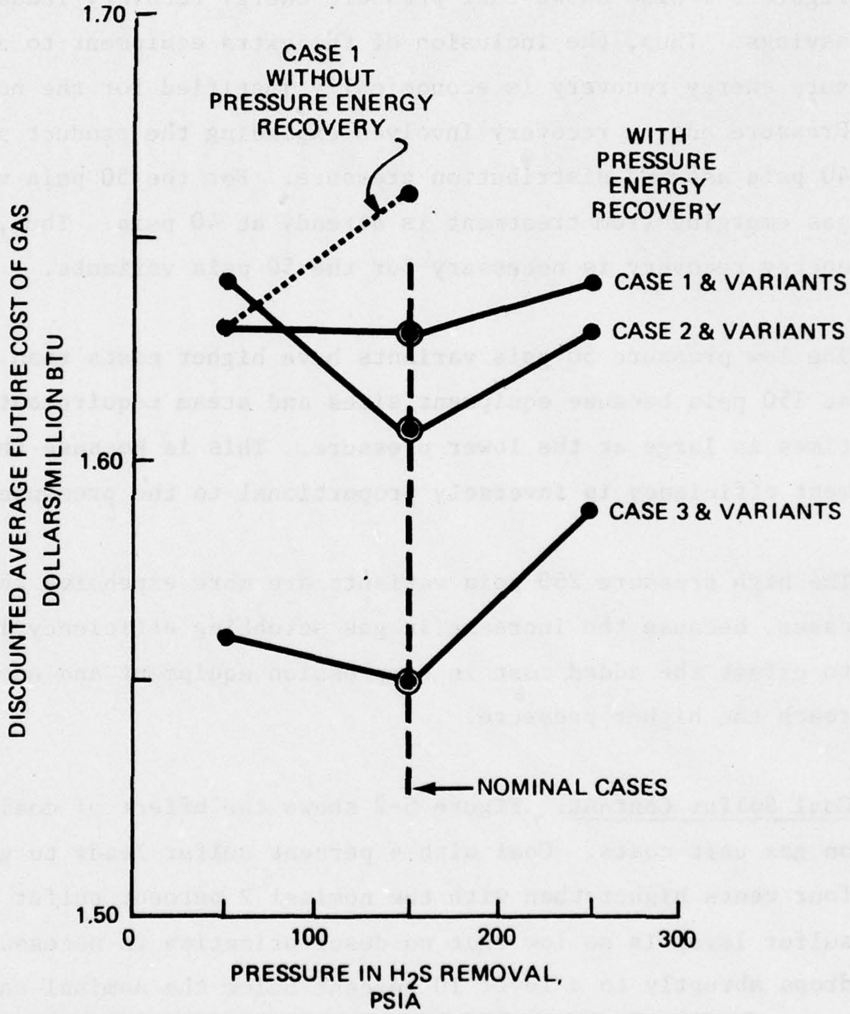
The cost summaries for the three nominal cases and 18 variants are tabulated in Appendix F. The product unit costs are shown in plots in this subsection.

Nominal Conditions. Parameter settings for the nominal cases are as follows:

Gas treatment pressure:	150 psia
Pressure energy recovery:	Included
Feed coal sulfur content:	2%
Lb SO_2 emitted/10^6 Btu coal:	1.2

The variants were generated by deviating from the nominal with one parameter at a time.

Gas Treatment Pressure and Pressure Energy Recovery. Figure 5-1 shows the effect of gas treatment pressure on gas cost. For all three cases,



CASE:	CASE 1	CASE 2	CASE 3	UNITS
BLAST MODE	O_2/STEAM	O_2/STEAM	AIR/STEAM	
GASIFIER	ENTD SOLIDS	FLUID SOLIDS	FLUID SOLIDS	
SULFUR/COAL	2	2	2	%
SO_2/COAL	1.2	1.2	1.2	LB $SO_2/10^6$ BTU HHV
TRTG PRESS.	150	150	150	PSIA

Figure 5-1. EFFECT OF GAS TREATMENT PRESSURE ON GAS COST

the 150 psia nominal pressure setting gave lower costs than the variants at 50 psia and 250 psia. Thus, the nominal cases are at a gas treatment pressure that is close to optimum.

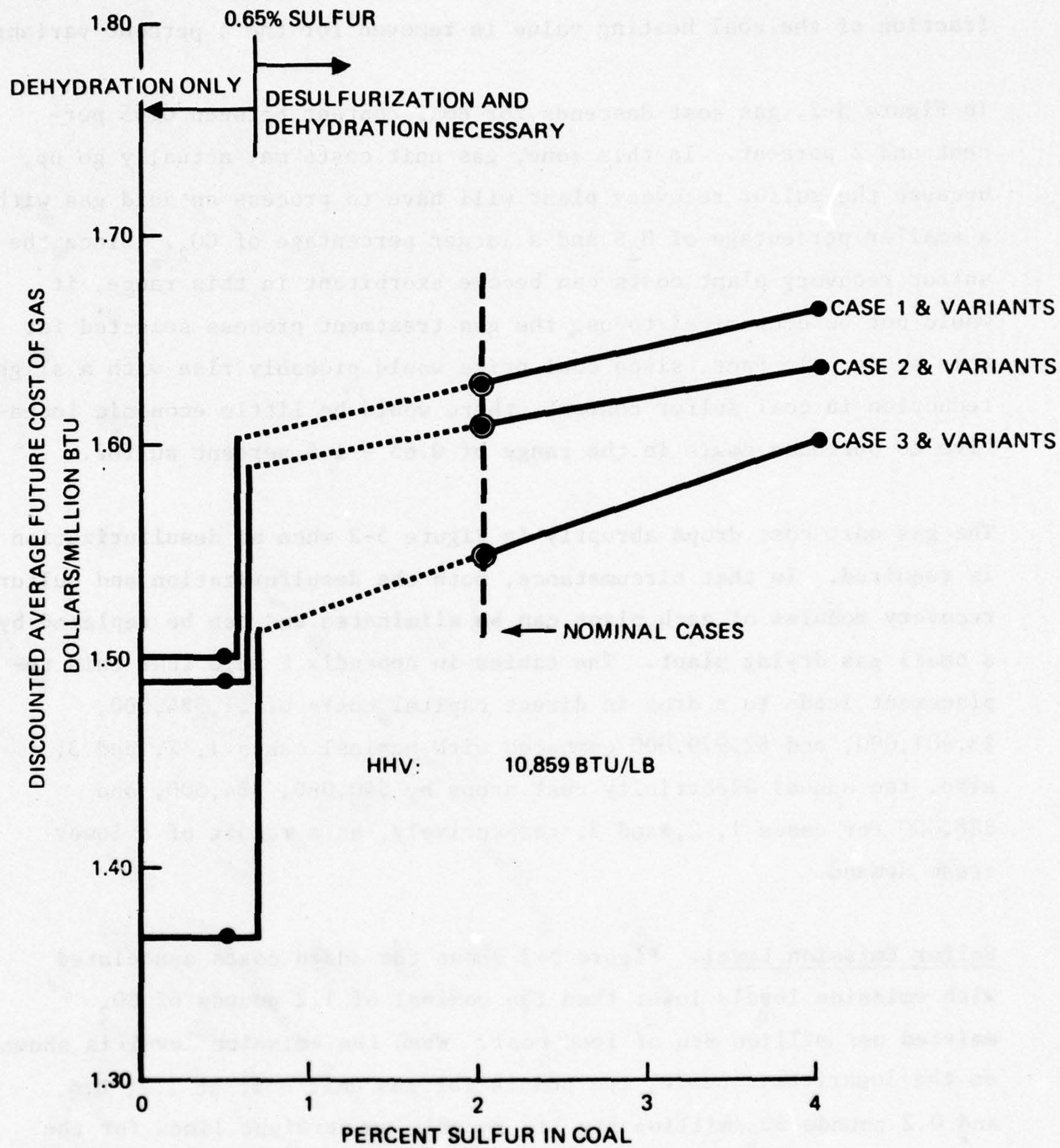
Figure 5-1 also shows that pressure energy recovery leads to significant savings. Thus, the inclusion of the extra equipment to accomplish pressure energy recovery is economically justified for the nominal cases. Pressure energy recovery involves expanding the product gas down to the 40 psia assumed distribution pressure. For the 50 psia variants product gas emerging from treatment is already at 40 psia. Thus, no pressure energy recovery is necessary for the 50 psia variants.

The low pressure 50 psia variants have higher costs than the nominals at 150 psia because equipment sizes and steam requirements are three times as large at the lower pressure. This is because the gas treatment efficiency is inversely proportional to the pressure of treatment.

The high pressure 250 psia variants are more expensive than the nominal cases, because the increase in gas scrubbing efficiency is not enough to offset the added cost in compression equipment and energy needed to reach the higher pressure.

Coal Sulfur Content. Figure 5-2 shows the effect of coal sulfur content on gas unit costs. Coal with 4 percent sulfur leads to gas costs two to four cents higher than with the nominal 2 percent sulfur cases. When the sulfur level is so low that no desulfurization is necessary, the gas cost drops abruptly to a level 10 percent below the nominal cases. This assumes that the coal price is still only 25 dollars per ton when the sulfur level is at one-half a percent or less.

The gas unit cost is higher for the 4 percent sulfur variant, because the capacity of the sulfur recovery plant must double, and the desulfurization plant equipment and steam requirements go up slightly. Also,



CASE:	CASE 1	CASE 2	CASE 3	UNITS
BLAST MODE	O_2/STEAM	O_2/STEAM	AIR/STEAM	
GASIFIER	ENTD SOLIDS	FLUID SOLIDS	FLUID SOLIDS	
SULFUR/COAL	2	2	2	%
SO_2/COAL	1.2	1.2	1.2	LB $SO_2/10^6$ BTU HHV
TRTG PRESS.	150	150	150	PSIA

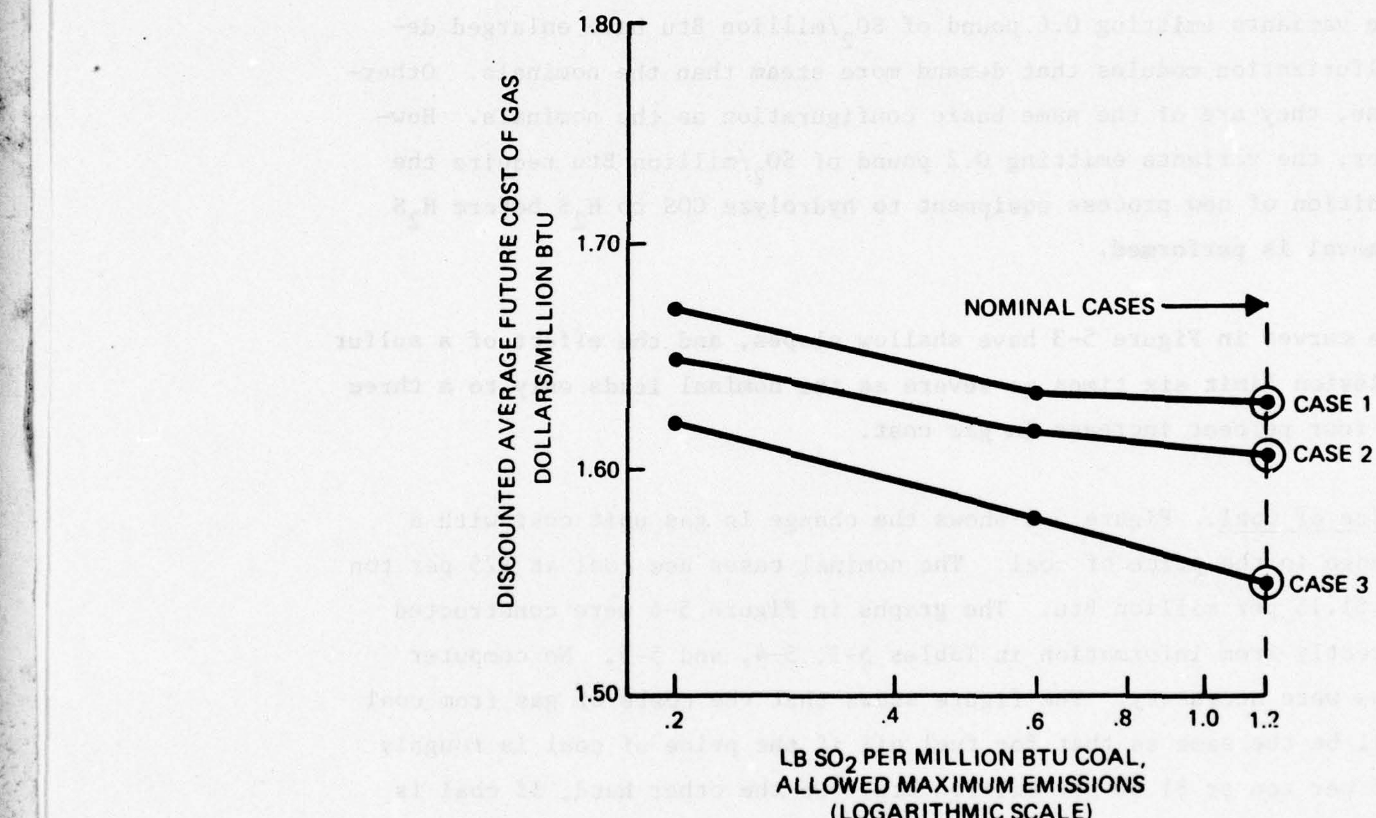
Figure 5-2. EFFECT OF COAL SULFUR ON GAS COST

since sulfur contributes part of the feed coal heating value, a larger fraction of the coal heating value is removed for the 4 percent variants.

In Figure 5-2, gas cost descends for coal content between 0.65 percent and 2 percent. In this zone, gas unit costs may actually go up, because the sulfur recovery plant will have to process an acid gas with a smaller percentage of H_2S and a larger percentage of CO_2. Since the sulfur recovery plant costs can become exorbitant in this range, it would not be economical to use the gas treatment process selected for this study. In fact, since coal price would probably rise with a slight reduction in coal sulfur content, there would be little economic incentive to purchase coals in the range of 0.65 - 1.5 percent sulfur.

The gas unit cost drops abruptly in Figure 5-2 when no desulfurization is required. In that circumstance, both the desulfurization and sulfur recovery modules of each plant can be eliminated and can be replaced by a small gas drying plant. The tables in Appendix F show that this replacement leads to a drop in direct capital costs of $1,984,000, $3,401,000, and $2,979,000 compared with nominal cases 1, 2, and 3; also, the annual electricity cost drops by $40,000, $84,000, and $88,000 for cases 1, 2, and 3, respectively, as a result of a lower steam demand.

Sulfur Emission Level. Figure 5-3 shows the added costs associated with emission levels lower than the nominal of 1.2 pounds of SO_2 emitted per million Btu of feed coal. When the emission level is shown on the logarithmic scale, the points for gas unit cost at 1.2, 0.6, and 0.2 pounds SO_2/million Btu lie roughly on straight lines for the three cases. Such a logarithmic dependance is expected when measuring cost versus severity of separation. As an illustration, note that the sulfur recovery section includes two sections, a Claus plant and a Scott plant. The Claus plant removes 95 percent of the sulfur present in the acid gas fed to it. Five percent of the sulfur leaves with the



CASE:	CASE 1	CASE 2	CASE 3	UNITS
BLAST MODE	O_2/STEAM	O_2/STEAM	AIR/STEAM	
GASIFIER	ENTD SOLIDS	FLUID SOLIDS	FLUID SOLIDS	
SULFUR/COAL	2	2	2	%
SO_2/COAL	1.2	1.2	1.2	LB $SO_2/10^6$ BTU HHV
TRTG PRESS.	150	150	150	PSIA

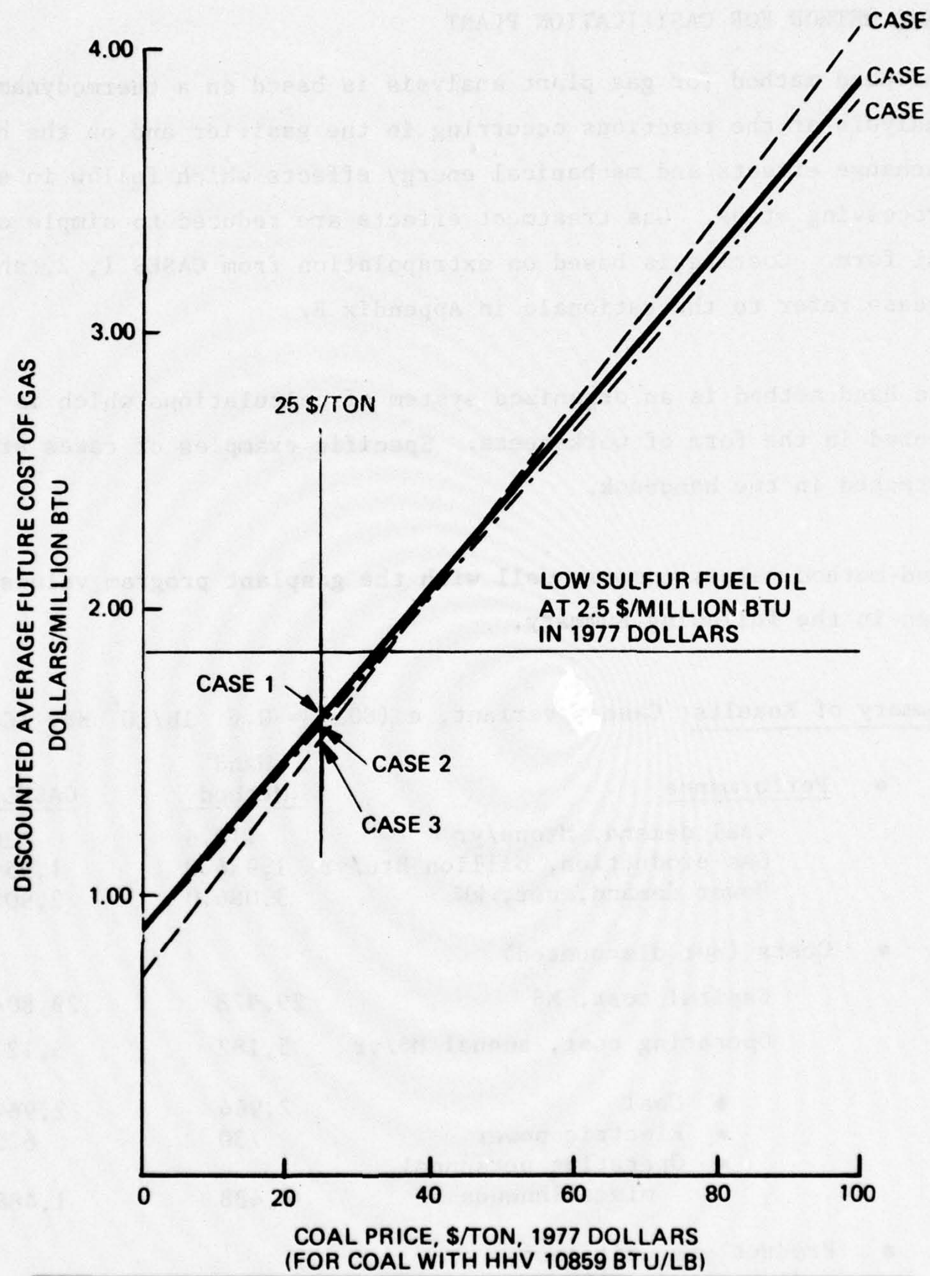
Figure 5-3. EFFECT OF EMISSION LEVEL ON GAS COST

tailgas. The Scott plant, in turn, treats the tailgas and removes 95 percent of the sulfur entering with the tailgas. The Scott plant, which reduced the sulfur level by the second factor of 20, costs approximately the same as the Claus plant, which reduced the sulfur content by the first factor of 20.

The variants emitting 0.6 pound of SO_2/million Btu have enlarged desulfurization modules that demand more steam than the nominals. Otherwise, they are of the same basic configuration as the nominals. However, the variants emitting 0.2 pound of SO_2/million Btu require the addition of new process equipment to hydrolyze COS to H_2S before H_2S removal is performed.

The curves in Figure 5-3 have shallow slopes, and the effect of a sulfur emission limit six times as severe as the nominal leads only to a three or four percent increase in gas cost.

Price of Coal. Figure 5-4 shows the change in gas unit cost with a change in the price of coal. The nominal cases use coal at $25 per ton or $1.15 per million Btu. The graphs in Figure 5-4 were constructed directly from information in Tables 5-3, 5-4, and 5-5. No computer runs were necessary. The figure shows that the costs of gas from coal will be the same as that for fuel oil if the price of coal is roughly $35 per ton or $1.60 per million Btu. On the other hand, if coal is available at $10 per ton or $0.46 per million Btu, the gas unit cost will be under $1.20 per million Btu.



CASE:	CASE 1	CASE 2	CASE 3	UNITS
BLAST MODE	O_2/STEAM	O_2/STEAM	AIR/STEAM	
GASIFIER	ENTD SOLIDS	FLUID SOLIDS	FLUID SOLIDS	
SULFUR/COAL	2	2	2	%
SO_2/COAL	1.2	1.2	1.2	LB $SO_2/10^6$ BTU HHV
TRTG PRESS.	150	150	150	PSIA

Figure 5-4. EFFECT OF COAL PRICE ON GAS COST

HAND METHOD FOR GASIFICATION PLANT

The hand method for gas plant analysis is based on a thermodynamic analysis of the reactions occurring in the gasifier and on the heat exchange effects and mechanical energy effects which follow in several processing steps. Gas treatment effects are reduced to simple empirical form. Costing is based on extrapolation from CASES 1, 2, and 3. Please refer to the rationale in Appendix B.

The hand method is an organized system of calculations which is presented in the form of worksheets. Specific examples of cases are attached in the handbook.

Hand method values compare well with the gasplant program values, as seen in the following summary.

Summary of Results; Case 1 variant, e $(SO_2) = 0.6$ lb/10^6 HHV (Coal)

	Hand Method	GASPLANT
• Performance		
Coal demand, Mtons/yr	118.6	120.7
Gas production, billion Btu/yr	1,936.0	1,959.4
Power demand, net, kW	3,086.0	2,900.0
• Costs (not discounted)		
Capital cost, M$	29,978	29,804
Operating cost, annual M$/yr	5,182	5,121
• Coal	2,964	2,969
• Electric power	730	673
• Operating personnel, miscellaneous	1,488	1,488
• Product gas unit cost, discounted, $$10^6$ Btu	1.67	1.63

BOILER DERATING

The results of the analysis technique developed in this study are detailed in the Handbook computation sheets, where examples are shown, the essential results from the computations of fuel change effects on the performance of three boiler types are listed in Table 5-6.

Apparent trends are discussed for each boiler type.

Oil-Fueled Water Tube Boiler. Conversion of this type of boiler to burn coal-derived gas fuels results in a reduction of combustion temperature which is caused by the reduced heating values of the coal-derived gases. This results in lower heat transfer performance throughout the boiler, thereby reducing the boiler efficiency.

The reduced heating value of the coal-derived gas fuels also requires a significant increase in fuel flow into the boiler furnace to achieve the same total fuel heat release rate, requiring an increase in combustion products gas flow through the boiler. To avoid major modifications to the boiler and its combustion air supply system, total increase in combustion product gas flow is limited to 1.10 times the original fuel case. This would cause a further reduction in steam capacity in addition to that caused by the combustion temperature reduction.

Coal-Fueled Water Tube Boiler. This type of boiler was originally designed to burn stoker-fed coal with 40 percent excess air at a combustion temperature of about 2300 F. Converting this type of boiler to coal-derived fuel gases from oxygen-blown conversion processes (286 and 269 Btu/cu ft), results in an increase in the combustion temperature and in the fuel heat release rate. This causes an increase in steam capacity and boiler efficiency. Burning fuel gas from the air blown conversion process (125 Btu/cu ft) results in a lower combustion temperature and a higher combustion products gas flow rate than for the original coal-fueled case. This causes a small reduction in steam

Table 5-6

ESTIMATED PERFORMANCE EFFECTS OF FUEL CHANGE IN THREE TYPES OF BOILERS

	Original Fuel		Conversion Gas Fuel		Performance Factors		
Boiler Type	Type	Combustion Temperature (F)	High Heat Value (Btu/cu ft)	Combustion Temperature (F)	Combustion Gas Flow	Steam Capacity	Efficiency
Water Tube Boiler	No.6	3980	286	2950	1.10	.74	.92
Maximum steam capacity	Oil	3980	269	2710	1.10	.67	.91
of 200,000 lb/hr.		3980	125	2170	1.10	.51	.87
Water Tube Boiler	Stoker	2300	286	2950	1.00	1.35	1.05
Maximum Steam Capacity	Coal	2300	269	2710	1.00	1.22	1.03
of 200,000 lb/hr.		2300	125	2170	1.10	.97	.94
Fire Tube Boiler	No. 6	3980	286	2950	1.10	.77	.95
Maximum Steam Capacity	Oil	3980	269	2710	1.10	.70	.94
of 10,000 lb/hr.		3980	125	2170	1.10	.53	.90

capacity and a reduction in boiler efficiency.

Oil-Fueled Fire Tube Boiler. Conversion of this type of boiler to coal-derived fuel gas results in a decrease in steam capacity and boiler efficiency similar to those for the oil-fueled water tube boiler type. Restriction of the combustion products gas flow rate to 1.10 times the original rate further reduces the steam capacity.

In summary, the estimated performance effects support the following two general conclusions regarding conversion of the subject boilers to burn coal-derived fuel gases:

- Either water tube or fire tube boilers designed originally to burn fuel oil will suffer significant derating unless major modifications are made to the boilers and their combustion air supply systems.

- Boilers designed originally for stoker-fed coal fuel will sustain or improve their rating without having to undergo major modifications.

Section 6

CONCLUSIONS

TASK ITEMS

A study of coal gasification for the Navy has led to the conceptual designs of three plants producing fuel gas, representing three combinations of reactor type and blast mode. The design is the result of tradeoff consideration.

The gasification section is optimized in the sense of operating in the range of conditions proper to the reactor types. The gas treatment section operates at a pressure demonstrated to be optimum.

The objective of providing the Navy with a method of estimating cost of major plant components is met with a procedure which accounts for all the parameters cited as variable. The method displays process rates as well as plant components costs. It yields finally the unit product cost, according to the Navy costing procedures.

A boiler derating procedure is also provided.

Description of the systems and results of analyses are presented in a final report. The procedures noted above are presented separately in a handbook.

ECONOMIC RESULTS

Based on the discounted outlays for fuels bought or produced in the future, the cost of fuel gas from the above plants is cheaper than

fuel oil: \$1.55 - \$1.64 per 10^6 Btu versus \$1.85 per 10^6 Btu. These costs are based on two-percent sulfur in a bituminous coal at \$25/ton and on an emissions limit of 1.2 lb SO_2 per 10^6 Btu (coal).

An expected downtrend in gas cost with reducing sulfur content in coal is found. The trend reflects a saving of one or two cents per 10^6 Btu for energy sulfur reduction of one-percent.

The cost of sulfur cleanup rises at accelerating rate as the emissions limit approaches zero. The rise amounts to about 2 or 3 cents per 10^6 Btu for every 50 percent reduction in SO_2 limit.

The recovery of power from pressurized fuel gas emerging from gas treatment results in a significant saving of about 5 cents per 10^6 Btu.

Section 7

REFERENCES

"Steam - Its Generation and Use," Babcock & Wilcox, New York, N.Y., 1972.

J.K. Salisbury, Power Volume, Section 7, Kent's Mechanical Engineering Handbook, 12th edition, John Wiley, New York, N.Y., 1954.

H.C. Hottel and A.F. Sarofim, Radiative Transfer, McGraw-Hill Book Company, New York, N.Y., 1967.

Appendix A

GASPLANT COMPUTER PROGRAM

The Coal Technology Department of Bechtel Corporation's Research and Engineering has developed a computer program for modeling coal gasification plant stream flows, utilities, equipment sizing, and capital costs. The computer routine, called Program GASPLANT was built to perform conceptual design calculations. Program GASPLANT produces a readable five-part readout dealing with the gasifier, main plant streams, utility streams, the check on mass and heat balances, and costs.

The program calculates the composition and emergence temperature of the gas produced in high temperature gasification. This calculation takes into account the thermodynamic equilibrium between CO, H_2, CO_2, H_2O, and CH_4 at the exit temperature predicted by a rigorous heat balance. Special condition settings appropriate to commercial entrained bed and fluidized bed gasifiers are incorporated.

The program explicitly calculates the composition and total flow of each of 55 streams in a representative H_2S scrubbing and sulfur recovery complex. The H_2S removal module uses equilibrium data on M-Pyrol to describe the simultaneous absorption of H_2S, COS, CO_2, and H_2O and subsequent processing in a module containing four separation towers, 11 heat exchangers and a two-stage compressor. Heat and material balances for a Claus sulfur recovery plant and a Scott tailgas treating plant are calculated.

The program also has a special utility subroutine that automatically sizes cooling tower's, boiler feed water treatment facilities, and

cooling water and condensate pumps and decides whether auxiliary boilers are needed. The utility routine totals the steam, electricity, cooling, and plant fuel services needed by a particular plant and tabulates the heat and mass flow to each plant unit needing service.

A cost package is included in the program that produces component cost details and totals for capital and operating costs in the same form as generated by Bechtel cost engineers. Costs of 16 common plant equipment and plant module types are generated automatically in terms of capacity and other relevant parameters.

A special subroutine has been built to check and assure mass and energy balances over each plant unit. Program GASPLANT places at the programmer's disposal an intelligible collection of Fortran subroutine calls for describing a particular plant. This exclusive Bechtel methodology has been given the name FLOWLANG. FLOWLANG plant descriptions are unusually easy to follow because each 29 allowed chemical species has its chemical formula or a common name as its global Fortran variable name. For instance, CO_2, designates the molar flow of carbon dioxide in a stream under consideration. This methodology has been used to create a description of the three gasification plants considered in the present study.

The most advanced version of Program GASPLANT is available on Bechtel's UNIVAC 1110 computer system. The Fortran coding occupies 70,000 words of disc storage, and the compiled program occupies 60,000 words of core on the Univac 1110. A run takes less than a minute. Program GASPLANT is ideal for optimization studies in which the least cost configuration is sought, and for design studies requiring frequent updating of the flow diagram.

Appendix B

RATIONALE FOR HAND METHOD

INTRODUCTION

A short method of plant design and costing is presented as the means of rationalizing the effects of several parameters in coal gasification. The parameters are comprised of the important properties of the feedstock and of the the product stream. The resultant effects are capital costs of plant modules and unit cost of product.

The problem to be faced is that of finding the plant module sizes and the operating requirements sufficient to support costing. The choice of method is narrowed by the need to assure a system that will operate in a practical way; the gas composition and reactor temperature should be reasonably known, for example. The approach is to observe that most of the parameters can be studied on a thermodynamic basis, accepting a few simplicities. The remainder, comprised of sulfur input and sulfur emissions, can be evaluated only by complex analysis. A precalculation of such effects can be reduced to empirical adjustments for these parameters.

This approach supports both the sizing of modules and the estimation of operating needs.

Costing then follows by adjusting the module costs obtained separately for the nominal plant. The latter values are the result of the detailed computer analysis.

DISCUSSION

Parametric Effects

The treatment of the following parameters is needed:

- Characteristics of Coal
  - Type
  - Heating Value
  - Moisture Content
  - Ash Content
  - Sulfur Content
- Pressure of the gas treatment section
- Heating value of gas
- Emission limit on sulfur in gas

The following effects are to be derived:

- Performance of the plant
- Capital cost of the modules comprising the plant
- Cost of the product gas

Rationale for Analysis

Gasification. The gasification of coal is analyzed by deriving the stoichiometric process that occur in the gasifier and then finding the heat effects. The energy effects in gasification are complemented by those in the final full combustion of gas. Two steps lead to final products of combustion:

(1) The partial combustion of coal that results from the addition of enough oxygen to bring essentially all the carbon into the gaseous state

(2) The combustion of the gas resulting as a fuel, liberating about three quarters of the heating, value originally available in the coal.

The sum of the above heat effects is the overall heat effect in full combustion of coal. The heat effect of Step (1); the partial combustion

of coal, is about one-quarter that of the overall combustion, generally. If it is possible to estimate the effects in either (1) or (2), or both, then the basic energy effects are fixed.

The heat effects for both processes are derived in the hand method. The combustibles are assumed to be carbon monoxide and hydrogen only. The distribution between these components is a function of the reactor temperature and the relative amounts of steam, oxygen, and carbon charged, characteristic of the type of reactor and the blast mode. Thus, the fraction of gasified carbon, that becomes carbon monoxide is reasonably known. Its value is not critical to finding the heat effects, but it supports a rough estimation of gas composition. This point is based on the following:

> Let $n_o(C)$ be the mols of carbon gasified, comprised of CO and CO_2 only; CH_4 is negligible. Oxygen, available from decomposition of some water and also directly from the gasifier blast, can be found after reaction in the carbon monoxide and dioxide:
>
> $$2n_o(O_2,\ \text{blst}) + n_2(H_2O_d) = n_2(CO) + n_2(CO_2) \text{ atoms oxygen}$$
>
> Now:
>
> $$n_2(CO) = \alpha \cdot n_o(C)$$
>
> $$n_2(CO_2) = (1 - \alpha) \cdot n_o(C)$$

Substitution of these terms in the balance leads by rearrangement to:

$$\alpha n_o(C) + n_2(H_2O_d) = 2\, n_o(C) - 2\, n_1(O_2,\ \text{blst})$$

Addition of net hydrogen from coal to both sides produces the total combustibles, if sulfur is ignored for the moment:

$$\alpha n_o(C) + n_2(H_2O_d) + n_2(H_2\ \text{net}) = 2\, n_o(C)$$

$$-2n_1(O_2,\ \text{blst}) + n_2(H_2,\ \text{net})$$

The terms on the left comprise the CO and H_2, expressed as the easily found sum on the right. Addition of the terms for CO_2, H_2O (undecomposed), and miscellaneous components yields total mols of products:

$$n_o(C) + n_2(H_2, \text{net}) + n_1(H_2O, \text{blst}) +$$
$$n_2(N_2) + n_o(S).$$

These terms are easily found as reactants charged to the gasifier.

The above shows that the amount of combustibles formed,

$$2n_o(C) - 2n_1(O_2, \text{blst}) + n_2(H_2, \text{net}),$$

is independent of α. The high heating value of the gas is nearly independent also, because the unit heating values for hydrogen and carbon monoxide are about equal.

Steam Generation. The rate of steam generation is found from the rate of delivery of sensible heat to the waste heat recovery section of the gasifier module.

Reactor Temperature. A rough check on the reactor temperature is performed to assure slagging or non-slagging conditions.

Energy Management. The compression of gas to the nominal 150 pisa gas treating plant pressure plus the compression of air for delivery to the gasifier directly or to the oxygen plant for liquefaction is evaluated for energy consumption. Energy recovery from steam, after subtraction of some energy diversion to the gas treatment plant is estimated. Any deficiency in energy is made up by purchase of electricity.

Development of the foregoing results, after defining the performance of the gasifier is performed by conventional methods. Some details are omitted here but are apparent from inspection of the computations sheets.

Costing. The costing of modules is performed by adjusting the corresponding costs in the nominal plant, based on changes in heat effects, gas rate, or sulfur rates.

Nomenclature

Variables.

n (i) = Number of mols of component i.

α = Fraction of carbon in coal converted to gas.

Subscripts. A number of the formal chemical symbols for single molecular components are used, as with H_2O, CO_2, C, O_2, N_2, and S. In addition, some are augmented as follows:

H_2O_d = Water which undergoes decomposition.

O_2, blst = Free oxygen in blast.

H_2, net = Net hydrogen available from coal.

Numbers denote certain stream identities:

0 = Coal feed.

1 = Blast stream entering the gasifier.

2 = Hot raw reactor gases.

Appendix C

BOILER DERATING METHOD

This section defines the first-order theoretical effects of the temperature and mass flow changes on the boiler performance ratings and suggests certain limitations to fuel changes that can be accepted by a boiler. It is assumed in the subsequent evaluations that fuel changes are acceptable only if minor changes to the boiler are required. Therefore, boiler tubing, air blower capacities, furnace dimensions, and most boiler auxiliaries are assumed to remain unchanged. Only those physical changes required to inject the low-Btu gas are considered.

Combustion Temperature Effects

Theoretical adiabatic equilibrium reaction temperatures for combustion of several fuels in air are plotted as a function of percent excess air in Figure C-1. This figure presents calculated results for the following fuels:

- No. 2 diesel oil
- No. 6 residual oil
- Natural gas
- Stoker-fed coal from the Unit Mine
- Coal-derived fuel gases: 286 Btu/ft^3, 269 Btu/ft^3, 157 Btu/ft^3, 125 Btu/ft^3,

Each fuel is burned in excess air. Conventional boiler design practice has established the excess air range for each fuel to be as shown on Figure C-1 (Ref. 7-2). Boilers designed for natural gas fuel use 5 to 10 percent excess air; oil fuel designs use 8 to 15 percent; stoker-fed coal designs use 30 to 50 percent; and coal-derived gas designs are expected to use 7 to 12 percent.

The computed adiabatic temperatures of Figure C-1 are valuable for comparison of potential combustion temperatures of the various fuels at various excess air combustion conditions. Relative changes in these potential combustion temperatures from a reference to a new operating condition are subsequently used to estimate corresponding changes in boiler operating performance.

Concerning Radiant Heat Transfer, it is assumed that heat transfer to water walls in the boiler furnace is predominantly by radiation. Then, defining the effects of fuel changes on radiant heat transfer also defines the performance changes in the boiler furnace. The basic equation for radiant heat transfer is:

$$q = \sigma eA(T_1^4 - T_2^4), \text{ (Ref. 7-1)} \qquad \text{(C-1)}$$

where,

q = heat transfer rate, Btu/sec.
σ = Stefan-Boltzman constant.
A = effective heat transfer area, ft^2.
T_1 = temperature of combustion gas, R.
T_2 = temperature of water walls, R.
e = effective emissivity of combustion gas.

Variables remaining constant in the application of Equation C-1 to this gas radiation situation are:

- σ, defined as a constant
- A will not change, since the furnace walls remain unchanged in switching from oil to gas fuel
- T_2, temperature of water walls will remain essentially constant due to high boiling heat transfer rate

Identifying the reference heat transfer use by subscript 1 (prior to fuel change) and the new heat transfer case by subscript 2 (after change to low-Btu gas fuel), the ratio of new to reference case heat transfer reduces to:

$$\frac{q_2}{q_1} = \frac{\sigma_2}{\sigma_1} \times \frac{e_2}{e_1} \times \frac{A_2}{A_1} \times \frac{(T_1^4 - T_2^4)_2}{(T_1^4 - T_2^4)_1} \qquad \text{(C-2)}$$

This reduces to the following, σ and A being constant:

$$\frac{q_2}{q_1} = \frac{e_2}{e_1} \times \frac{(T_1^4 - T_2^4)_2}{(T_1^4 - T_2^4)1} \qquad \text{(C-3)}$$

Equation (C-3) indicates that the rate of heat transfer in Case 2 relative to that in Case 1 is known, if the corresponding ratios of emissivity and the temperature function can be defined.

The temperature function $T_1^4 - T_2^4$ was computed for T_2 equal to 826 R and 798 R, with T_1 variable from T_2 up to 5,000 R. Figure C-2 shows the results as a plot of the temperature function variation with T_1. This defines the temperature function for use in Equation (C-3), with T_1 values obtainable from Figure C-1 for particular fuels and excess air combinations.

Defining accurately the effective emissivities for radiation heat transfer from combustion gases to the water walls is difficult. Emissivity is a function of the partial pressure of each radiating gas in the mixture, effective mean beam length to the absorbing surface, and temperature of radiating gas mixture (Ref. 7-3). In this study, the only radiating combustion product gases are CO_2 and H_2O.

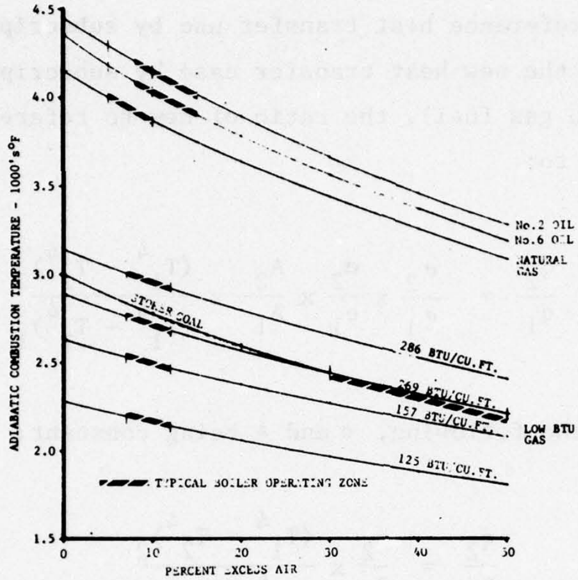


Figure C-1. Theoretical Adiabatic Combustion Temperature Variation with Percent Excess Air for Various Boiler Fuels

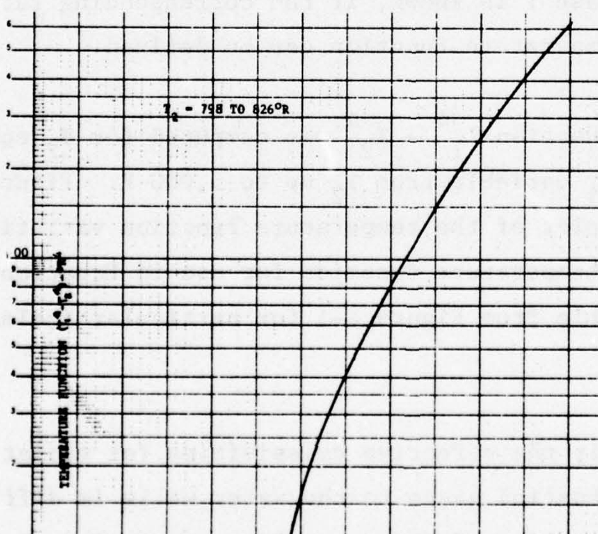


Figure C-2. Radiant Heat Transfer Function Variation with Combustion Reaction Temperature

The diatomic gases, nitrogen and oxygen , are transparent and do not contribute to radiation. Flames with ash content are more luminous and must have a small factor added to the computed emissivity. Oil fuel flame emissivities are increased by adding 0.05. Coal fuel flame emissivities are increased by 0.10 (Ref. 7-3, p. 488). A constant mean beam length of 10 feet is used to calculate relative emissivities with gas partial pressures taken from the computer computations for adiabatic combustion analyses used for Figure C-1. Computed emissivities are shown in Figure C-3.

Mass Flow Effects

Increases in mass flow through the boilers, resulting from increased fuel flow when changing to coal-derived gas, affects both radiation and convection heat transfer. Combustion product gas velocities within the fixed geometry boilers must increase with increasing gas flow rates. This reduces the stay time within the furnace section of a boiler, thus reducing the time period for radiant heat transfer. In the convection tube section of a boiler, increased velocity also increases the heat transfer rate. The net effect may be to either decrease or increase total heat transfer to the steam, depending on initial heat transfer conditions.

Table C-1 lists the results of the computed combustion product gas quantities. The last column in the table lists the pounds of combustion product gas produced per 10,000 Btu of fuel energy release, for each fuel. Note that the combustion gas quantity increases as the fuel heating value decreases. It is assumed that combustion product gas flows resulting from a fuel change are limited to a maximum 10 percent increase.

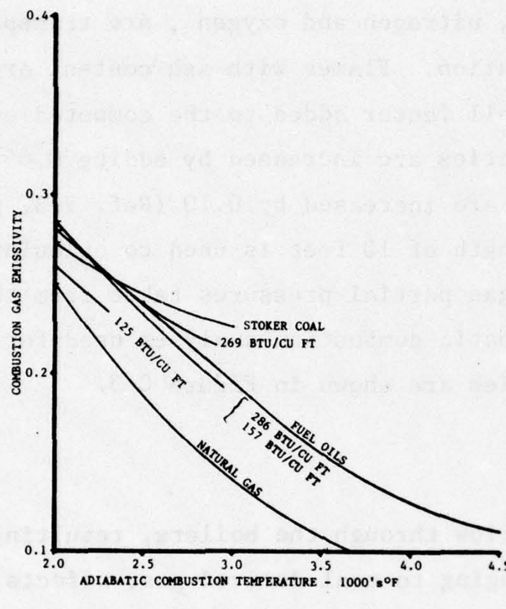


Figure C-3. Combustion Product Gas Emissivity Variation with Adiabatic Combustion Temperature

Table C-1

FUEL IN AIR COMBUSTION PRODUCT GAS QUANTITIES

Fuel	Stoichiometric Air/Fuel Ratio	Excess Air (%)	Actual Air/Fuel Ratio	High Heating Value (Btu/lb)	Combustion Gas (lb/lb fuel)	Products Flow (lb/10,000 Btu)
No. 2 Diesel Oil	13.051	12	14.617	19,600	15.617	7.968
No. 6 Residual Oil	12.845	12	14.386	18,600	15.386	8.272
Natural Gas	16.434	8	17.749	23,250	18.749	8.064
Coal, Unit Mine	11.545	40	16.163	10,709	17.163	16.03
Low-Btu Gas, 286 Btu/cu ft	4.6154	10	5.0769	5,280	6.0769	11.51
Low-Btu Gas, 269 Btu/cu ft	4.9349	10	5.4284	5,314	6.4284	12.10
Low-Btu Gas, 157 Btu/cu ft	2.3730	10	2.6103	2,708	3.6103	13.33
Low-Btu Gas, 125 Btu/cu ft	2.3977	10	2.6375	1,921	3.6375	18.94

Radiant Heat Transfer, the transfer rate is not affected by gas flow changes. However, the total heat radiated from a pound of gas in the furnace portion of a boiler is a direct function of the stay time in the furnace. Shorter stay time results in less heat transferred. The effect of stay time is not linear, since the heat transfer rate at combustion temperature (time zero) is much greater than when existing from the furnace at a lower temperature. The effect is estimated in the following sample solution.

Assume furnace conditions of 3,900^{o}F combustion temperature and 2,000^{o}F at exit from the furnace. Since the radiant heat transfer rate is predominantly a function of temperature, the ratio of the rate at exit to the rate at combustion is:

$$\frac{q_2}{q_1} = \frac{(T_1^{\,4} - T_2^{\,4})_2}{(T_1^{\,4} - T_2^{\,4})_1} = \frac{38}{365} = 0.104,$$

where the temperature function $(T_1^{\,4} - T_2^{\,4})$ is read from Figure C-2 for the furnace exit temperature (subscript 2) and the combustion temperature (subscript 1.) Since the stay time changes affect only the furnace exit conditions, a change of 10 percent in the stay time causes a change in total heat transfer of approximately:

$$\frac{\text{Stay Time}}{\text{Stay Time}} \times \frac{q_2}{q_1} = 0.1 \times 0.104 = 0.104$$

Therefore, if stay time changes by less than 10 percent, its effect on radiant heat transfer can be assumed negligible.

Convection Heat Transfer. Heat transfer from combustion product gases to steam in the tubular convection section of a boiler is resisted by external and internal tube surface convection coefficients and a tube wall conduction coefficient. For high conductivity steel tubing with relatively thin walls, the wall conduction resistance is negligibly low, and the overall heat transfer coefficient becomes:

$$U = \frac{U_o U_i}{U_o + U_i}, \quad \text{(Ref. 7-1)} \qquad \text{(C-4)}$$

where,

U_o = outer tube wall heat transfer coefficient, Btu/ft^2-hr-^{o}F.
U_i = inner tube wall heat transfer coefficient, Btu/ft^2-hr-^{o}F.

Equation C-4 in terms of heat transfer resistance is:

$$R = R_o + R_i, \quad \text{(Ref. 7-1)} \qquad \text{(C-5)}$$

where,

$$R = \frac{1}{U} = \text{ft}^2\text{-hr }^o\text{F/Btu} \qquad \text{(C-6)}$$

The overall heat transfer coefficient, Equation (C-4), changes as the inner and outer tube surface heat transfer coefficients change. If this effect is examined for the convection sections for the boilers in this study, the effect of changes in the heat transfer coefficient on one surface while the other remains constant is easily illustrated. Figure C-4 shows this effect for (U_o/U_i) as a variable condition. A 10 percent change in the outer surface heat transfer coefficient results in about a 5 percent change in the overall coefficient.

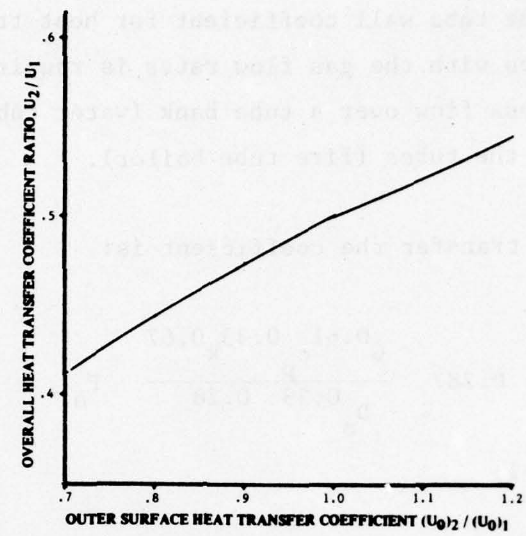


Figure C-4. Overall Convection Heat Transfer Coefficient Variation with that of Outer Tube Surface

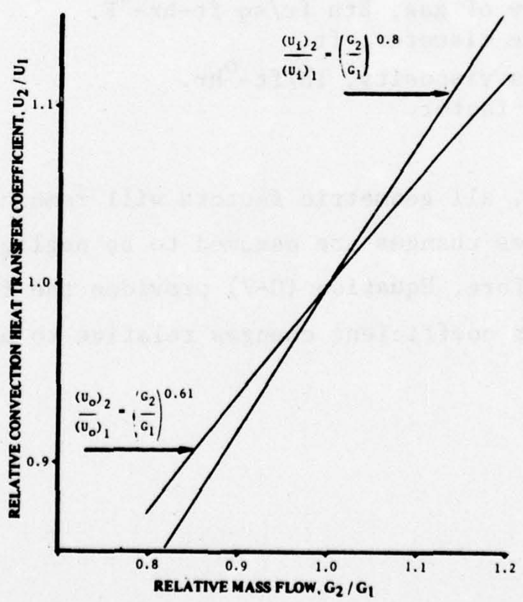


Figure C-5. Relative Convection Heat Transfer Coefficient Variation with Relative Combustion Products Gas Mass Flow

The variations of the tube wall coefficient for heat transfer from combustion product gases with the gas flow rates is required for the following — for gas cross flow over a tube bank (water tube boiler) and for gas flow within the tubes (fire tube boiler).

For cross flow heat transfer the coefficient is:

$$U_o = 0.287 \frac{G^{0.61} c_p^{0.33} k^{0.67}}{D_o^{0.39} u^{0.28}} F_a \qquad \text{(C-7)}$$

(Ref. 7-1)

where,

- G = mass velocity, or mass flow of gas over the tubes, lb/hr sq-ft of cross-sectional area.
- c_p = specific heat at constant pressure, Btu/lb-^oF.
- k = conductivity of gas, Btu ft/sq ft-hr-^oF.
- D_o = outside tube diameter, ft.
- u = absolute gas viscosity, lb/ft-^{o}hr.
- F_a = arrangement factor.

For use in this study, all geometric factors will remain constant, and gas physical properties changes are assumed to be negligible for first order effects. Therefore, Equation (C-7) provides the following relationship for heat transfer coefficient changes relative to mass flow changes:

$$\frac{(U_o)_2}{(U_o)_1} = \left(\frac{G_2}{G_1}\right)^{0.61} \qquad \text{(C-8)}$$

For turbulent gas flow within the tubes, the heat transfer coefficient is

$$U_i = 0.023 \left[\frac{G^{0.8} c_p^{0.4} k^{0.6} T_b^{0.8}}{D_i^{0.2} u^{0.4} T_f^{0.8}}\right] \qquad \text{(C-9)}$$

(Ref. 2-6)

Where,

T_b = average bulk absolute temperature of gas, R

T_f = average film absolute temperature, R

Again, in this study, all geometric and physical property changes are neglected for first order effects analysis. Equation 9 provides the following for heat transfer coefficient changes relative to mass flow changes:

$$\frac{(U_i)_2}{(U_i)_1} = \left(\frac{G_2}{G_1}\right)^{0.8} \qquad \text{(C-10)}$$

The total heat transferred from a pound of gas in the convection section of a boiler is also a direct function of the stay time in that section. Shorter stay time results in less heat transferred. The effect of stay time is not linear, since the heat transfer rate at combustion product temperature entering the convection section (time zero) is much greater than when exiting from the convection section at a lower temperature. The effect is estimated in the following sample solution.

Assume the convection section conditions of $2000^{o}F$ at entry and $466^{o}F$ at exit. Since the convection heat transfer rate will vary predominantly as a function of the temperature difference, the ratio of the rate at exit to the rate at entry is:

$$\frac{q_2}{q_1} = \frac{(T_2 - T_s)}{(T_1 - T_s)} = \frac{466\text{-}366\ F}{2{,}000\text{-}366\ F} = \frac{100}{1{,}634} = 0.0061,$$

where,

T_2 = convection section exit temperature = $366^{o}F$.

T_1 = convection section entry temperature = $2000^{o}F$.

T_s = steam temperature = 366 F.

Since the stay time changes affect only the convection section exit conditions, a change of 10 percent in the stay time causes a change in total heat transfer of approximately:

$$\frac{\Delta \text{ Stay Time}}{\text{Stay Time}} \times \frac{q_2}{q_1} = 0.10 \times 0.061 = 0.0061$$

Therefore, if stay time changes by less than 10 percent, its effect on convection heat transfer can be assumed to be negligible.

To estimate the boiler performance rating factors for conversion to cool-derived gas fuel, it is necessary to systematically apply the analytical methods discussed in this section. The computational methodology and work sheets are defined in the handbook volume of this report.

Appendix D

STREAM COMPOSITIONS AND FLOW DATA

Table D-1

GASIFICATION PLANT CASE 1
STREAM COMPOSITION AND FLOW RATES

	Stream No.	010-100 Raw Coal	100-300 Sized Dry Coal	300-310 Hot Gas	310-320 Cool Gas	320-400 Particle Free Cool Gas	400-500 Compressed Cool Gas	500-855 Clean Indst'l Gas	860-100 F.G. to Coal Prep	855-860 F.G. to F.G. Header	855-920 F.G. to Market	850-200 O_2 PLT Intake Air	290-300 Oxygen Blast
MMOLS/HR	Comp												
	C	–	–	0.040	0.004	–	–	–	–	–	–	–	–
	H_2	–	–	0.714	0.714	0.714	0.714	0.712	0.004	0.007	0.705	–	–
	O_2	–	–	–	–	–	–	–	–	–	–	0.648	0.648
	N_2	–	–	0.028	0.028	0.028	0.028	0.028	–	–	0.028	2.438	0.013
	S	–	–	–	–	–	–	–	–	–	–	–	–
	ASH	–	–	0.056	0.005	–	–	–	–	–	–	–	–
	H_2O	–	–	0.496	0.496	0.070	0.014	–	–	–	–	–	–
	CO	–	–	1.297	1.297	1.297	1.297	1.296	0.006	0.012	1.284	–	–
	CO_2	–	–	0.144	0.144	0.144	0.144	0.121	0.001	0.001	0.120	–	–
	CH_4	–	–	–	–	–	–	–	–	–	–	–	–
	H_2S	–	–	0.018	0.018	0.018	0.018	0.005	–	–	0.005	–	–
	COS	–	–	0.001	0.001	0.001	0.001	0.001	–	–	0.001	–	–
	TOTAL	–	–	2.794	2.707	2.272	2.216	2.163	0.001	0.020	2.143	3.086	0.661
MLB/HR	Comp												
	C	18.2	18.2	0.5	0.1	–	–	–	–	–	–	–	–
	H_2	1.1	1.1	1.4	1.4	1.4	1.4	1.435	0.007	0.013	1.421	–	–
	O_2	1.8	1.8	–	–	–	–	–	–	–	–	20.7	20.7
	N_2	0.4	0.4	0.8	0.8	0.8	0.8	0.8	–	–	0.80	68.3	0.4
	S	0.6	0.6	–	–	–	–	–	–	–	–	–	–
	ASH	6.5	6.5	3.4	0.3	–	–	–	–	–	–	–	–
	H_2O	1.5	0.6	8.9	8.9	1.3	0.3	–	–	–	–	–	–
	CO	–	–	36.3	36.3	36.3	36.3	36.304	0.181	0.338	35.966	–	–
	CO_2	–	–	6.3	6.3	6.3	6.3	5.3	0.026	0.050	5.27	–	–
	CH_4	–	–	–	–	–	–	–	–	–	–	–	–
	H_2S	–	–	0.6	0.6	0.6	0.6	0.2	–	–	0.2	–	–
	COS	–	–	0.1	0.1	0.1	0.1	0.1	–	–	0.1	–	–
	TOTAL	30.1	29.2	58.4	54.9	46.8	45.8	44.1	0.215	0.401	43.7	89.0	21.1
MSCFH	Gas	–	–	1060.3	1027.3	862.2	841.0	821.0	4.2	7.7	813.2	1171.1	250.8
°F	Temp	77	212	2350	350	104	100	100	100	100	100	80	220
PSIA	Press	14.7	14.7	35	35	165	40	40	40	40	40	35	35
BTU/DSCF, (BTU/LB)HHV		–	(11047)	294.5	298.8	299.3	299.3	300.7	300.7	300.7	300.7	–	–
MMBTU/HR, Sensible Ht		–	1.371	51.9	5.4	0.4	0.4	0.4	–	–	–	0.1	0.7

$M = 10^3$
$MM = 10^6$

Case 1:
Gasifier Type: Entrained Solids
Reactor Mode: Oxygen/Steam Blast
Sulfur/Coal: 2 %
SO_2/Coal: 1.2 lb $SO_2/10^6$ Btu HHV

Table D-1 (Continued)

	Stream Description	200-870 Vent Nitrogen	280-300 Blast Steam	300-330 Waste Solids from Gasifier	810-310 Waste Heat BLR Feed H_2O	310-820 Hi-Press Steam	320-880 Venturi Cond. Wst. H_2O	500-600 Acid Gas	860-600 Tail Gas Red. Fuel	600-930 Sulfur Product	330-940 Wst. Solids to Disposal	310-330 Wst. Solids from WHR	320-330 Waste Solids from Dust Removal	840-320 Water for Dust Removal
MMOLS/HR	Comp													
	C	–	–	–	–	–	–	–	–	–	–	–	–	–
	H_2	–	–	–	–	–	–	–	0.003	–	–	–	–	–
	O_2	–	–	–	–	–	–	–	–	–	–	–	–	–
	N_2	2.422	–	–	–	–	–	–	–	–	–	–	–	–
	S	–	–	–	–	–	–	–	–	–	–	–	–	–
	ASH	–	–	–	–	–	–	–	–	–	–	–	–	–
	H_2O	–	0.363	–	1.983	1.922	–	0.001	–	–	–	–	–	–
	CO	–	–	–	–	–	–	–	0.006	–	–	–	–	–
	CO_2	–	–	–	–	–	–	0.022	–	–	–	–	–	–
	CH_4	–	–	–	–	–	–	–	–	–	–	–	–	–
	H_2S	–	–	–	–	–	–	0.013	–	–	–	–	–	–
	COS	–	–	–	–	–	–	–	–	–	–	–	–	–
	TOTAL	2.422	0.363	–	1.983	1.922	–	0.036	0.009	–	–	–	–	–
MLB/HR	Comp													
	C	–	–	0.4	–	–	–	–	–	–	0.9	0.4	0.1	–
	H_2	–	–	–	–	–	–	–	0.006	–	–	–	–	–
	O_2	–	–	–	–	–	–	–	–	–	–	–	–	–
	N_2	67.8	–	–	–	–	–	–	–	–	–	–	–	–
	S	–	–	–	–	–	–	–	–	0.4	–	–	–	–
	ASH	–	–	3.0	–	–	–	–	–	–	6.5	3.0	0.5	–
	H_2O	–	6.5	–	35.7	34.6	7.6	0.02	–	–	1.1	–	1.1	535.7
	CO	–	–	–	–	–	–	–	0.157	–	–	–	–	–
	CO_2	–	–	–	–	–	–	1.0	0.024	–	–	–	–	–
	CH_4	–	–	–	–	–	–	–	–	–	–	–	–	–
	H_2S	–	–	–	–	–	–	0.43	–	–	–	–	–	–
	COS	–	–	–	–	–	–	–	–	–	–	–	–	–
	TOTAL	67.8	6.5	3.4	35.7	34.6	7.6	1.45	0.187	0.4	8.5	3.4	1.7	535.7
MSCFH	Gas	919.1	–	–	–	–	–	13.5	3.5	–	–	–	–	–
°F	Temp	80	275	77	366	900	125	100	100	300	77	77	77	75
PSIA	Press	14.7	45.0	14.7	1200	1055	14.7	30	40	14.7	14.7	14.7	14.7	14.7
BTU/DSCF, (BTU/LB)HHV		–	–	–	–	–	–	234.4	300.7	–	–	–	–	–
MMBTU/HR, Sensible Ht.		0.0	0.6	–	–	–	–	–	–	–	–	–	–	–

Table D-2

GASIFICATION PLANT CASE 2
STREAM COMPOSITION AND FLOW RATES

	Stream No. Description	010-100 Raw Coal	100-300 Sized Dry Coal	300-310 Hot Gas	310-320 Cool Gas	320-400 Particle Free Cool Gas	400-500 Compressed Cool Gas	500-855 Clean Indst'l Gas	860-600 Tail Gas Red. Fuel	855-860 F.G. to F.G. Header	855-920 F.G. to Market	850-200 O_2 Plant Intake Air	290-300 Oxygen Blast
MMOLS/HR	Comp												
	C	–	–	.084	0.008	–	–	–	–	–	–	–	–
	H_2	–	–	1.042	1.042	1.042	1.042	0.966	0.007	0.007	0.959	–	–
	O_2	–	–	–	–	–	–	–	–	–	–	0.472	0.472
	N_2	–	–	0.024	0.024	0.024	0.024	0.025	–	–	0.025	1.775	0.010
	S	–	–	–	–	–	–	–	–	–	–	–	–
	ASH	–	–	0.054	0.005	–	–	–	–	–	–	–	–
	H_2O	–	–	0.826	0.826	0.075	0.014	–	–	–	–	–	–
	CO	–	–	0.839	0.839	0.839	0.839	0.784	0.006	0.006	0.778	–	–
	CO_2	–	–	0.409	0.409	0.409	0.409	0.369	0.003	0.003	0.366	–	–
	CH_4	–	–	0.040	0.040	0.040	0.040	0.082	0.001	0.001	0.081	–	–
	H_2S	–	–	0.015	0.015	0.015	0.015	0.003	–	–	0.003	–	–
	COS	–	–	0.003	0.003	0.003	0.003	0.002	–	–	0.002	–	–
	TOTAL	–	–	3.336	3.211	2.447	2.386	2.231	0.017	0.017	2.214	2.247	0.482
MLB/HR	Comp												
	C	17.5	17.5	1.0	0.1	–	–	–	–	–	–	–	–
	H_2	1.1	1.1	2.1	2.1	2.1	2.1	1.95	0.02	0.02	1.93	–	–
	O_2	1.7	1.7	–	–	–	–	–	–	–	–	15.1	15.1
	N_2	0.4	0.4	0.7	0.7	0.7	0.7	0.69	–	–	0.69	49.7	0.3
	S	0.6	0.6	–	–	–	–	–	–	–	–	–	–
	ASH	6.2	6.2	3.2	0.3	–	–	–	–	–	–	–	–
	H_2O	1.4	1.4	14.9	14.9	1.4	0.3	–	–	–	–	–	–
	CO	–	–	23.5	23.5	23.5	23.5	21.96	0.170	0.170	21.79	–	–
	CO_2	–	–	18.0	18.0	18.0	18.0	16.25	0.13	0.13	16.12	–	–
	CH_4	–	–	0.6	0.6	0.6	0.6	1.31	0.01	0.01	1.30	–	–
	H_2S	–	–	0.5	0.5	0.5	0.5	0.11	–	–	0.11	–	–
	COS	–	–	0.2	0.2	0.2	0.2	0.11	–	–	0.11	–	–
	TOTAL	28.9	28.9	64.7	60.9	47.0	45.9	42.39	0.33	0.33	42.06	64.8	15.4
MSCFH	Gas	–	–	1266.0	1218.6	928.6	905.5	846.6	6.6	6.6	840.0	852.7	182.9
°F	Temp	77	77	1800	350	104	100	100	100	100	100	80	68
PSIA	Press	14.7	14.7	55	55	35	165	40	40	40	40	55	55
BTU/DSCF, (BTU/LB) HHV		(10709)	(10709)	277.4	277.7	277.7	277.7	291.6	291.6	291.6	291.6	–	–
MMBTU/HR, Sensible Heat		–	–	49.1	6.7	0.5	0.4	0.4	–	–	–	–	–

$M = 10^3$
$MM = 10^6$

Case 2:

Gasifier Type:	Fluidized Bed
Reactor Mode:	Oxygen/steam blast
Sulfur/Coal:	2 %
SO_2/Coal:	1.2 lb $SO_2/10^6$ Btu coal HHV

Table D-2 (Continued)

Stream No. Description	200-870 Vent Nitrogen	280-300 Blast Steam	300-330 Waste Solids from Gasifier	810-310 Waste Heat BLR Feed H_2O	310-820 Hi-Press Steam	320-880 Venturi Cond. Wst. H_2O	600-930 Sulfur Product	310-330 Wst. Solids from WHR	320-330 Waste Solids from Dust Removal	330-940 Wst. Solids to Disposal	840-320 Water for Dust Removal	500-600 Acid Gas
MMOLS/HR Comp												
C	—	—	—	—	—	—	—	—	—	—	—	—
H_2	—	—	—	—	—	—	—	—	—	—	—	—
O_2	—	—	—	—	—	—	—	—	—	—	—	—
N_2	1.765	—	—	—	—	—	—	—	—	—	—	—
S	—	—	—	—	—	—	—	—	—	—	—	—
ASH	—	—	—	—	—	—	—	—	—	—	—	—
H_2O	—	1.355	—	1.589	1.539	—	—	—	—	—	—	0.003
CO	—	—	—	—	—	—	—	—	—	—	—	—
CO_2	—	—	—	—	—	—	—	—	—	—	—	0.076
CH_4	—	—	—	—	—	—	—	—	—	—	—	—
H_2S	—	—	—	—	—	—	—	—	—	—	—	0.012
COS	—	—	—	—	—	—	—	—	—	—	—	0.001
TOTAL	1.765	1.355	—	1.589	1.539	—	—	—	—	—	—	0.072
MLB/HR Comp												
C	—	—	0.9	—	—	—	—	0.9	0.1	1.9	—	—
H_2	—	—	—	—	—	—	—	—	—	—	—	—
O_2	—	—	—	—	—	—	—	—	—	—	—	—
N_2	49.4	—	—	—	—	—	—	—	—	—	—	—
S	—	—	—	—	—	—	0.4	—	—	—	—	—
ASH	—	—	2.94	—	—	—	—	2.94	0.3	6.2	—	—
H_2O	—	24.4	—	28.6	27.7	13.5	—	—	1.2	1.2	948.7	0.05
CO	—	—	—	—	—	—	—	—	—	—	—	—
CO_2	—	—	—	—	—	—	—	—	—	—	—	3.33
CH_4	—	—	—	—	—	—	—	—	—	—	—	—
H_2S	—	—	—	—	—	—	—	—	—	—	—	0.4
COS	—	—	—	—	—	—	—	—	—	—	—	0.1
TOTAL	49.4	24.4	3.84	28.6	27.7	13.5	0.4	3.84	1.6	9.3	948.7	3.0
MSCFH Gas	669.8	—	—	—	—	—	—	—	—	—	—	27.3
°F Temp	80	900	77	230	900	125	300	77	77	77	75	100
PSIA Press	14.7	65	14.7	1200	1055	14.7	14.7	14.7	14.7	14.7	14.7	30
BTU/DSCF,(BTU/LB) HHV	—	—	—	—	—	—	—	—	—	—	—	$116.\overline{9}$
MMBTU/HR, Sensible Heat	—	9.5	—	—	—	13.00	.007	—	—	—	—	

Table D-3

GASIFICATION PLANT CASE 3
STREAM COMPOSITION AND FLOW RATES

	Stream Description	010-100 Raw Coal	100-300 Sized Dry Coal	300-310 Hot Gas	310-320 Cool Gas	320-400 Particle Free Cool Gas	400-500 Compressed Cool Gas	500-855 Clean Indst'l Gas	860-600 Tail Gas Red. Fuel	855-860 F.G. to F.G. Header	855-920 F.G. to Market	850-200 Air Comp. Intake Air	290-300 Air Blast
MMOLS/HR	Comp												
	C	–	–	0.098	0.009	–	–	–	–	–	–	–	–
	H_2	–	–	0.784	0.784	0.784	0.784	0.792	0.002	0.006	0.786	–	–
	O_2	–	–	–	–	–	–	–	–	–	–	0.739	0.739
	N_2	–	–	2.796	2.796	2.796	2.796	2.792	0.008	0.021	2.771	2.779	2.779
	S	–	–	–	–	–	–	–	–	–	–	–	–
	ASH	–	–	0.063	0.006	–	–	–	–	–	–	–	–
	H_2O	–	–	0.430	0.430	0.161	0.030	0.001	–	–	–	–	–
	CO	–	–	1.224	1.224	1.224	1.224	1.196	0.004	0.009	1.187	–	–
	CO_2	–	–	0.281	0.281	0.281	0.281	0.256	0.001	0.002	0.254	–	–
	CH_4	–	–	0.001	0.001	0.001	0.001	0.005	–	–	0.005	–	–
	H_2S	–	–	0.018	0.018	0.018	0.018	0.004	–	–	0.004	–	–
	COS	–	–	0.003	0.003	0.003	0.003	0.002	–	–	0.002	–	–
	TOTAL	–	–	5.698	5.552	5.268	5.137	5.047	0.015	0.038	5.047	3.518	3.518
MLB/HR	Comp												
	C	20.4	20.4	1.2	0.1	–	–	–	–	–	–	–	–
	H_2	1.2	1.2	1.6	1.6	1.6	1.6	1.60	0.004	0.01	1.59	–	–
	O_2	2.0	2.0	–	–	–	–	–	–	–	–	23.6	23.6
	N_2	0.5	0.5	76.3	78.3	78.3	78.3	78.23	0.224	0.59	77.64	77.9	77.9
	S	0.7	0.7			–	–	–	–	–	–	–	–
	ASH	7.2	7.2	3.8	0.4	–	–	–	–	–	–	–	–
	H_2O	1.7	1.7	7.8	7.8	2.9	0.5	–	–	–	–	–	–
	CO	–	–	34.3	34.3	34.3	34.3	0.01	0.112	0.25	33.24	–	–
	CO_2	–	–	12.4	12.4	12.4	12.4	11.26	0.044	0.08	11.18	–	–
	CH_4	–	–	–	–	–	–	0.08	–	–	0.08	–	–
	H_2S	–	–	0.6	0.6	0.6	0.6	0.13	–	–	0.13	–	–
	COS	–	–	0.2	0.2	0.2	0.2	0.12	–	–	0.120	–	–
	TOTAL	33.7	33.7	140.1	135.6	130.3	127.9	124.91	0.384	0.93	123.98	101.5	101.5
MSCFH	Gas	–	–	2162.4	2107.0	1999.2	1949.5	1915.5	5.700	14.3	1901.2	1335.1	1335.1
°F	Temp	77	77	2090.5	350	104	100	100	100	100	100	80	483
PSIA	Press	14.7	14.7	55	55	35	165	40	165	40	40	55	–
BTU/DSCF,(BTU/LB)HHV		(10709)	(10709)	133.7	129.8	129.4	129.4	128.6	128.9	128.6	128.6	–	–
MMBTU/HR, Sensible Ht		–	–	91.4	10.9	1.0	0.8	0.8	–	–	–	0.1	10.1

$M = 10^3$
$MM = 10^6$

Case 3:
Gasifier Type: Fluidized Bed
Reactor Mode: Air/Steam Blast
Sulfur/Coal: 2 %
SO_2/Coal: 1.2 lb $SO_2/10^6$ Btu HHV

Table D-3 (Continued)

	Stream No. Description	280-300 Blast Steam	300-330 Waste Solids from Gasifier	310-330 Waste Solids from WHR	810-310 Waste Heat BLR Feed Water	310-820 Hi-Press Steam	320-880 Venturi Cond. Waste H_2O	600-930 Sulfur Product	320-330 Waste Solids from Dust Removal	840-320 Waste Solids to Disposal	840-320 Water for Dust Removal	500-600 Acid Gas
MMOLS/HR	Comp											
	C	–	–	–	–	–	–	–	–	–	–	–
	H_2	–	–	–	–	–	–	–	–	–	–	–
	O_2	–	–	–	–	–	–	–	–	–	–	–
	N_2	–	–	–	–	–	–	–	–	–	–	–
	S	–	–	–	–	–	–	–	–	–	–	–
	ASH	–	–	–	–	–	–	–	–	–	–	–
	H_2O	0.523	–	–	3.133	3.072	–	–	–	–	–	0.002
	CO	–	–	–	–	–	–	–	–	–	–	–
	CO_2	–	–	–	–	–	–	–	–	–	–	0.047
	CH_4	–	–	–	–	–	–	–	–	–	–	–
	H_2S	–	–	–	–	–	–	–	–	–	–	0.014
	COS	–	–	–	–	–	–	–	–	–	–	0.001
	TOTAL	0.523	–	–	3.133	3.072	–	–	–	–	–	0.064
MLB/HR	Comp											
	C	–	1.0	1.0	–	–	–	–	0.2	2.2	–	–
	H_2	–	–	–	–	–	–	–	–	–	–	–
	O_2	–	–	–	–	–	–	–	–	–	–	–
	N_2	–	–	–	–	–	–	–	–	–	–	–
	S	–	–	–	–	–	–	0.5	–	–	–	–
	ASH	–	3.4	3.4	–	–	–	–	0.4	7.2	–	–
	H_2O	9.4	–	–	56.4	55.3	4.9	–	1.4	1.4	339.7	0.032
	CO	–	–	–	–	–	–	–	–	–	–	–
	CO_2	–	–	–	–	–	–	–	–	–	–	1.725
	CH_4	–	–	–	–	–	–	–	–	–	–	–
	H_2S	–	–	–	–	–	–	–	–	–	–	0.472
	COS	–	–	–	–	–	–	–	–	–	–	0.068
	TOTAL	9.4	4.4	4.4	56.4	55.3	4.9	0.5	2.0	10.8	339.7	2.298
MSCFH	Gas	–	–	–	–	–	–	–	–	–	–	21.3
°F	Temp	900	77	77	230	900	125	300	77	77	75	100
PSIA	Press	65	14.7	14.7	1200	1055	14.7	14.7	14.7	14.7	14.7	30
BTU/DSCF,(BTU/LB) HHV		–	–	–	–	–	–	–	–	–	–	176.2
MMBTU/HR, Sensible Heat		3.7	–	–	–	–	.087	.008	–	–	–	–

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Appendix E

EQUIPMENT LIST — CASES 1, 2, 3

Table E-1

EQUIPMENT LIST — CASES 1, 2, AND 3

	Case 1(1)		Case 2(2)		Case 3(3)	
	Quantity	Item	Quantity	Item	Quantity	Item
Section 100 Coal Handling						
Unloader	1	Receiving and transfer bin, 6,000 ft^3 capacity. Reinforced concrete, walls at 42° slope, 40' x 20' x 20' deep. Walls 2' thick. Concrete: 140 cy. Steel: 25,200 lb		Same as Column 1		Same as Column 1
Feeder	2	Feeder apron, 72" wide, 25 FPM (capacity: 750 TPH each) Cost: $6,000 each. Cost: $12,000				
Belt conveyor	1	Belt conveyor, 48" wide, 500 ft long L to L; equipped with tripper and symmetrical discharge chutes to both sides of belt. Elevation through 74' height. Cost: $250,000 each. Cost: $250,000				
Tunnel under stockpile	1	Tunnel under stockpile, 6' h x 6' w x 60' long, fed from 2 hoppers with walls at 45° incline. Reinforced concrete walls				
Feeder	2	Reciprocating feeder, 24" x 36". Capacity: 20 TPH, Total 40 TPH. Cost: $6,000 each. Cost $12,000				
Belt conveyor	1	Belt conveyor, 18" wide by 150' long; elevation through 30'. Cost: $55,000				
Surge bin	1	Surge bin: capacity, 125 ft^3 (5 minutes inventory), 5' x 5' x 5', discharge cone walls at 45° incline. Cost:				
Hammermill	1	Hammermill, Type 34A; capacity, 40 TPH; 99% solids reduction to minus 1/4 inch; with bars. Motor: 75 hp Cost: $10,000		No Item		No Item
Bulldozer	1	Bulldozer, wheeled, 4.0-4.5 C.Y. bucket, type 745-HB, Fiat-Allis, 240 hp max. Cost: $72,000		Same as Column 1		Same as Column 1
Belt conveyor	1	Belt conveyor, 18" wide x 250' long; elevation through 80'; capacity: 31 TPH. Cost: $56,000			1	Belt conveyor, 18" wide x 280' long; elevation through 90'; capacity: 36.5 TPH. Cost: $56,000

Notes: (1) Gasifier: entrained solids; oxygen/steam blast

(2) Gasifier: fluidized bed; oxygen steam blast

(3) Gasifier: fluidized bed; air/steam blast.

Table E-1 (Continued)

	Case 1[(1)] Quantity	Case 1[(1)] Item	Case 2[(2)] Quantity	Case 2[(2)] Item	Case 3[(3)] Quantity	Case 3[(3)] Item
Section 100 (Continued) **Coal Handling**						
Storage bin	1	Storage bin, circular; 1,100-ton capacity (coal); 30' ϕ x 84' high; reinforced concrete. Cost: $60,000		Same as Column 1		1,250-ton capacity (coal); 30' ϕ x 92' high; reinforced concrete.
Feeder, vibrating	1	Feeder, vibrating; 24" wide; capacity, 40 TPH; 0.8 kW. Cost: $5,000				
Conveyor, screw	1	Conveyor, screw; 9" ϕ 95% loading, 35 rpm. Cost: $8,000		No Item		No Item
Bowl mill	1	Bowl mill, 713 type, 900 rpm, 450 hp, to grind 70% of coal to minus 200 mesh, complete with air separator, fan, and ducts. Inlet gas temperature: 700 °F. Cost: $363,000.				
Day bin	1	Day bin (gasifier charge bin, and included with the system erected by gasifier vendor). Cost: part of gasifier module				
Section 200						
Oxygen Plant	1 1 1	Air compressor Air separator (cold box) Oxygen compressor, O_2 capacity: 250 TPD; 98% purity; 55 psia discharge	1 1 1	Air compressor Air separator (cold box) Oxygen compressor, O_2 capacity: 185 TPD; 98% purity; 55 psia discharge		No Item
	1	O_2 storage (capacity, 125 tons)	1	O_2 storage (capacity, 93 tons)		
Air Compression		No Item		No Item	1	Two-stage centrifugal compression interstage and post cooling: 3300 BHP; suction: 14.7 psia and 77°F; discharge: 57 psia and 270°F; ICFM: 22,900
Air Compression Driver		Not Applicable		Not Applicable	1	Steam turbine, feed steam at 1,055 psia, 900°F, condenses 20.4 M #/hr steam at 3 psia, bleeds 12.0 M #/hr at 65 psia

Table E-1 (Continued)

Section 300	Case 1[1] Quantity	Case 1[1] Item	Case 2[2] Quantity	Case 2[2] Item	Case 3[3] Quantity	Case 3[3] Item
Coal Gasification (including waste heat recovery and dust removal from gases)	1	Gasifier, entrained solids type, integral with process train comprised of waste heat recovery and dust cleanup sections. Gasifier capacity is 6 x 10^9 Btu/D with unit coal.	1	Gasifier, 12' ϕ x 175' high (approx.), integral with a process train comprised of waste heat recovery and dust cleanup sections. Gasifier capacity is 6 x 10^9 Btu/D, with unit coal.		Section module is the same as for Case 2, except as noted in this column:
					1	Gasifier: 14'
	1	Waste heat recovery section, receiving hot raw gas from the reactor at 2000 °F or less, and discharging gas at 350 °F	1	Waste heat recovery section, receiving hot raw gas from the reactor at 2000 °F or less, and discharging gas at 350 °F	1	Waste heat recovery section, enlarged steam capacity, 2000 °F maximum inlet temperature, discharge at 350 °F
	1	Wash tower, cooler	1	Dust removal cyclone		
	1	Water spray scrubber	1	Venturi scrubber for final dust removal with water spray to 1 grain per 1,000 scf		
	1	Slag quency and removal system				
	1	Coal feed system, with motive nitrogen carrier (available from oxyben plant)	1	Spray tower, for gas cooling		Same type of units as in Column 2, but larger gas flow.
			1	System for collection, grinding, cooling, and delivering of ash to holding bin for retransport		
			1	Settler tank for separation of residual dust from gas washwater		
			1	Lock hopper system for feeding coal (minus 1/4') to the gasifier vessel		Same units, but slightly larger coal feed capacity.
		Total Section Cost: $5,000,000		Total Section Cost: $5,000,000		Total Section Cost: $6,000,000
Section 400 Raw Gas Compression						
Driver	1	Steam turbine, feed steam at 1,055 psia, 900°F, condensing 16.7 M #/hr at 3 psia, and 6.5 M #/hr at 65 psia	1	Steam turbine, feed steam at 1,055 psia, 900°F, condensing 6.8 M #/hr at 3 psia, and 24.5 M #/hr at 65 psia	1	Steam turbine, feed steam at 1,055 psia, 900°F, condensing 37.1 M #/hr at 3 psia, and 9.4 M #/hr at 65 psia
	1	Vacuum condenser. Assumed: The condenser for the cold condensate is a specialized item available from the vendor of the above driver turbine. Condenses steam at 3 psia	1	Vacuum condenser. See note at Column 1. Condenses steam at 3 psia.	1	Vacuum condenser. See note at left. Condenses steam at 3 psia.
Gas Compression	1	Two-stage centrifugal compressor interstage and post cooling, 2,500 BHP; suction: 35 psia and 104°F; discharge: 167 psia and 288°F; ICFM: 6,650. Mol. wt.: 20.66	1	Two-stage centrifugal compressor interstage and post cooling, 2,700 BHP; suction: 35 psia and 104°F; discharge: 167 psia and 281°F; ICFM: 7,070. Mol. wt.: 19.2	1	Two-stage centrifugal compressor interstage and post cooling, 5,600 BHP; suction: 35 psia and 104°F; discharge: 167 psia and 287°F; ICFM: 15,200. Mol. wt.: 24.9

Table E-1 (Continued)

	Case 1(1) Quantity	Case 1(1) Item	Case 2(2) Quantity	Case 2(2) Item	Case 3(3) Quantity	Case 3(3) Item
Section 500						
Gas Treatment						
Heat exchangers	1	Interchanger, shell-and-tube, liquid-liquid, countercurrent, operating at 150 psig and 350°F, 16,930 ft², U = 80 Btu/ft² hr°F, mild steel	1	Interchanger, surface area: 17,800 ft²	1	Interchanger, surface area 32,000 ft²
	1	Heater, shell and tube, steam-liquid, operating at 150 psig and 350°F, U = 100 B/sfhf; exchange surface area: 757 ft² mild steel	1	Heater, surface area: 794 ft²	1	Heater, surface area: 1,430 ft²
	1	Cooler-condenser, shell-and-tube, liquid-vapor, mild steel, condensate withdrawal, surface: 82 ft², operating at 150 psig, 350°F maximum	1	Cooler-condenser, surface area: 86 ft²	1	Cooler-condenser, surface area: 154 ft²
	1	Kettle reboiler, shell-and-tube, steam/liquid-pyrol, 150 psig steam/8 psid vapor; area: 159 ft², mild steel	1	Kettle reboiler, surface area: 167 ft²	1	Kettle reboiler, surface area: 300 ft²
	1	Cooler, shell-and-tube, counterflow preferred, water-liquid pyrol; operating at 150 psig and 150°F maximum; area: 529 ft², mild steel	1	Cooler, surface area: 555 ft²	1	Cooler, surface area: 1,000 ft²
	1	Cooler-condenser, shell-and-tube, vapor/cooling water, operating at 8 psia, 350°F; water cooled, surface area: 80 ft², mild steel	1	Cooler-condenser, surface area: 84 ft²	1	Cooler-condenser, surface area: 150 ft²
	1	Cooler-condenser, shell-and-tube, vapor/cooling water, operating at 15 psig, 267°F, surface area: 6 ft² mild steel	1	Cooler-condenser, surface area: 6 ft²	1	Cooler-condenser, surface area: 10 ft²
	1	Cooler-condenser, shell-and-tube, vapor/cooling water, operating at 15 psig and 267/100°F, water condensate, surface area: 4 ft², mild steel	1	Cooler-condenser, surface area: 4 ft²	1	Cooler-condenser, surface area: 7 ft²
Scrub/Strip						
Tanks	3	Charcoal beds, vertical cylinders, 150 psia, 450°F operating, 4.25' 15' t-t, C-steel		Same as Column 1	3	Charcoal beds, vertical cylinders, 150 psia, 450°F operating, 6.22' x 15' t-t, C-steel
Charcoal charge	3	4,000 lb JxC, 416 Columbia activated carbon			3	8,700 lb JxC, 416 Columbia activated carbon
Exchanger, Heater	1	Steam heated, 425 psia, 450°F; fuel gas: 100°F-400°F, C-steel, shell-and-tube, 900 ft²			1	Steam heated, 425 psia, 450°F; fuel gas: 100°F-400°F, C-steel, shell-and-tube, 2,000 ft²
Exchanger, Cooler	1	Cooling water, 75-100°F; fuel gas: 400-100°F; C-steel, shell-and-tube, 150 psia, 400°F maximum, 1,250 ft²			1	Cooling water, 75-100°F; fuel gas: 400-100°F; C-steel, shell-and-tube, 150 psia, 400°F maximum, 2,750 ft²

Table E-1 (Continued)

	Case 1[1] Quantity	Case 1[1] Item	Case 2[2] Quantity	Case 2[2] Item	Case 3[3] Quantity	Case 3[3] Item
Section 500 (Continued) Gas Treatment						
Heat exchangers	1	Condenser, shell-and-tube, vapor/cooling water, operating at 8 psia and 225/100°F, water condensate, surface area: 85 ft^2, mild steel	1	Condenser, surface area: 89 ft^2	1	Condenser, surface area: 160 ft^2
	1	Kettle reboiler, steam heated, shell-and-tube, liquor at 225°F, 8 psia; surface area: 624 ft^2	1	Kettle reboiler, surface area: 655 ft^2	1	Condenser, surface area: 1,180 ft^2
	1	Heater (steam), shell-and-tube, liquor at 100/225°F, 20 psig, mild steel, surface area: 11 ft^2	1	Heater (steam), surface area: 11 ft^2	1	Heater (steam), surface area: 20 ft^2
Vessels	1	Tower, gas-scrubbing, 8 stages, 3.5'ϕ x 25', 150 psig at 110°F, mild steel	1	Tower, gas scrubbing, 4.1'ϕ x 25'	1	Tower, gas scrubbing, 5.5'ϕ x 25'
	1	Vessel, for flash and separation of supersaturated liquid feed, 4.5'ϕ x 20', 15 psig at 350°F, mild steel	1	Vessel, flash and separation, 4.5'ϕ x 20'	1	Vessel, flash and separation, 6'ϕ x 20'
	1	Tower, liquor stripping, 2 stages, 8 psia at 350°F, 1.5'ϕ x 15', mild steel Not Available	1	Tower, liquor stripping, 1.5'ϕ x 15'	1	Tower, liquor stripping, 2.0'ϕ x 15'
	1	Tower, fractionator, 4 stages, 8 psia at 350°F, 1.5'ϕ x 20', mild steel	1	Tower, fractionator, 1.5'ϕ x 20'	1	Tower, fractionator, 2.0'ϕ x 20'
Compressors	1	Compressor, reciprocating, in: 7.5 psia x 100°F x 66.4 scfm; out: 15.0 psia x 260°F; hp (brake): 4.8	1	Compressor, reciprocating, 70 scfm, 5.1 hp (brake)	1	Compressor, reciprocating, 125 scfm, 9.1 hp (brake)
	1	Compressor, reciprocating, in: 15 psia x 100°F x 66.4 scfm; out: 30 psia x 260°F; hp (brake): 4.8	1	Compressor, reciprocating, 65 scfm, 4.8 hp (brake)	1	Compressor, reciprocating, 118 scfm, 8.6 hp (brake)
Pumps	1	Pump, circulation of liquid pyrol, 810 gpm, 350°F, 80 hp brake	1	Pump, circulation of liquid, 840 gpm, 350°F, 83 hp brake	1	Pump, circulation of liquid, 1,150 gpm, 350°F, 150 hp brake
Section 600 Sulfur Recovery						
Claus plant	1	Feed gas: 13,440 scfh; vol % H_2S: 40.1%; mass rate: 5.5 TPD S	1	Feed gas: 27,270 scfh; vol % H_2S: 16.7%; mass rate 4.7 TPD S	1	Feed gas: 21,062 scfh; vol % H_2S: 25.6%; mass rate: 5.5 TPD S
Tail gas Treating plant	1	Tail gas: 20,000 scfh Vol % H_2S: 1.3%	1	Tail gas: 20,600 scfh Vol % H_2S: 0.8%	1	Tail gas: 25,000 scfh Vol % H_2S: 1.0%
		These plants are purchased from vendors as packages on a turn-key basis.				
Section 700 Scrub Liquor Dehydration				See Section 500		

Table E-1 (Continued)

	Case 1[1] Quantity	Case 1[1] Item	Case 2[2] Quantity	Case 2[2] Item	Case 3[3] Quantity	Case 3[3] Item
Section 800						
Auxiliaries	1	Water treatment facility for softening, for process purposes 0.080 x 10^6 lb/hr (160 gpm)	1	Water treatment facility for softening, for process purposes 0.140 x 10^6 lb/hr (280 gpm)	1	Water treatment facility for softening, for process purposes 0.10 x 10^6 lb/hr (200 gpm)
	1	Boiler feed water treating plant, converting softened water into deaerated, demineralized boiler feed, 0.070 x 10^6 lb/hr (140 gpm)	1	Boiler feed water treating plant, converting softened water into deaerated, demineralized boiler feed, 0.070 x 10^6 lb/hr (140 gpm)	1	Boiler feed water treating plant, converting softened water into deaerated, demineralized boiler feed, 0.085 x 10^6 lb/hr (170 gpm)
	1	Cooling tower, receives water at 100°F, discharges water at 75°F, at 3.9 x 10^6 lb/hr (7,800 gpm)	1	Cooling tower, receives water at 100°F, discharges water at 75°F, at 2.3 x 10^6 lb/hr (4,800 gpm)	1	Cooling tower, receives water at 100°F, discharges water at 75°F, at 3.9 x 10^6 lb/hr (7,800 gpm)
	1	Settler, receives dirty water 0.5 x 10^6 lb/hr (1,000 gpm), removes suspended mineral dust	1	Settler, receives dirty water 0.8 x 10^6 lb/hr (1,600 gpm), removes suspended mineral dust	1	Settler, receives dirty water 0.65 x 10^6 lb/hr (1,300 gpm), removes suspended mineral dust
Gas Expansion	1	Two-stage centrifugal expander, inlet: 150 psia, 280°F; outlet: 40 psia, 100°F; gas rate: 824 mscfh	1	Two-stage centrifugal expander, inlet: 150 psia, 280°F; outlet: 40 psia 100°F; gas rate: 877 mscfh	1	Two-stage centrifugal expander, inlet: 150 psia, 280°F; outlet: 40 psia, 100°F; gas rate: 1,920 msfch
Section 900						
Waste Disposal						
Solids storage bunker	1	Slag solids storage bunker, 5,000 ft^3, to receive solids from slag discharge section of Koppers Totzek gasifier; loaded with belt conveyor 30" wide by 200' long x 40' elevation	1	Char storage bunker, 5000 ft^3 receiving ash from char dust hopper (Section) by screw conveyor, discharging through chutes to slurry-char mixer (integral moistening) and thence to transport truck (see COSORB system)		Same as Column 2
Transport truck	1	Truck, 35 ton capacity, for slag solids; end-dump	1	Transport truck, 35 ton capacity, for solids; end-dump		
Screw conveyor		No Item	1	Screw conveyor, to handle ash fines at 40 lbs/cft, 3.5 TPH, elevation through 50'		
Barge loading facility — ash	1	Barge loading facility for disposal of slag solids at sea. Assumed available.	1	Barge loading for disposal of ash at sea; port facility for barge-loading of ash from dump assumed available.		
	1	Belt conveyor for transport of wet solids to holding/loading bin, 10 TPH capacity, 500 ft to 40 ft elevation		Same as Column 1		Same as Column 1
	1	Holding/loading bin (5,000 ft^2), 10' x 15' x 50' long, v-bottom, with 10 hp motor, closed top				
				Ash quench bin with spray array (6), v-bottom, with mixer and screw discharge features; bin dimensions: 8'ϕ x 10' deep, provision of cyclone separator at top to clean steam		Same as Column 2
Water treatment pond	1	Pond for chemical treatment of wastewater: softening and clarification; no organic wastes anticipated. Capacity:		Same as Column 1		Same as Column 1

Appendix F

SUMMARIZED INVESTMENT AND OPERATING COSTS FOR PARAMETRIC VARIANTS

Table F-1

CASE 1 AND VARIANTS
SUMMARIZED INVESTMENT AND OPERATING COSTS FOR VARIANTS
(Thousands of Dollars)

Run Number	3	6	9	12	15	18	21
Run Conditions	NOM	50 PSI	250 PSI	0.5% S	4% S	$e(SO_2)=.6$	$e(SO_2)=.2$
1977 Capital Costs							
100 Coal Preparation	2,850	2,850	2,850	2,850	2,850	2,850	2,850
200 Oxygen Supply	7,064	7,064	7,064	7,090	7,029	7,064	7,064
300 Gasification	5,426	5,426	5,426	5,426	5,426	5,426	5,476
400 Compression	1,706	514	2,068	1,713	1,700	1,706	1,700
500 Desulfurization Dehydration	889	1,785	744	210	950	955	1,477
600 Sulfur Recovery	1,332	1,332	1,332	—	1,822	1,348	1,408
700 Piping	1,825	1,825	1,825	1,825	1,825	1,825	1,825
800 Utilities	478	464	481	472	483	480	483
900 Waste Disposal	424	424	424	424	424	424	424
Total Direct Costs	21,994	21,684	22,214	20,010	22,509	22,078	22,657
Total Capital Costs	29,804	29,473	30,092	27,121	30,474	29,924	30,778
1977 Annual Costs							
Coal	2,969	2,969	2,969	2,989	2,943	2,969	2,969
Electricity	673	703	721	633	654	673	695
Operating Labor and Materials	1,488	1,488	1,488	1,098	1,488	1,488	1,488
Present Values 1977							
First Year Construction	4,306	4,259	4,348	3,919	4,403	4,324	4,448
Second Year Construction	7,829	7,741	7,904	7,124	8,005	7,860	8,084
Third Year Construction	10,685	10,566	10,788	9,723	10,925	10,728	11,034
Coal	36,424	36,424	36,424	36,669	36,105	36,424	36,424
Electricity	9,463	9,886	10,142	8,910	9,198	9,467	9,764
Operating Labor and Materials	9,679	9,679	9,679	7,142	9,679	9,679	9,679
Product Costs							
Total Project Present Value	78,386	78,558	79,285	73,487	78,314	78,482	79,433
Gas Production, 25 years (Billions of Btu)	48,118	48,192	48,205	49,031	47,100	48,017	47,627
Discounted Average Gas Cost ($ 1 million Btu)	1.63	1.63	1.64	1.50	1.66	1.63	1.67

Gasifier Type: Entrained Solids
Reactor Mode: Oxygen/Steam Blast

Table F-2

CASE 2 AND VARIANTS
SUMMARIZED INVESTMENT AND OPERATING COSTS FOR VARIANTS
(Thousands of Dollars)

Run Number	2	5	8	11	14	17	20
Run Conditions	NOM	50 PSI	250 PSI	0.5 %	4 % S	$e(SO_2)=.6$	$e(SO_2)=.2$
1977 Capital Costs							
100 Coal Preparation	2,228	2,228	2,228	2,228	2,228	2,228	2,228
200 Oxygen Supply	5,992	5,992	5,992	6,005	5,958	5,992	5,992
300 Gasification	6,016	6,016	6,016	6,016	6,016	6,016	6,016
400 Compression	1,758	530	2,143	1,767	1,750	1,758	1,747
500 Desulfurization Dehydration	904	1,851	786	217	1,186	1,215	1,538
600 Sulfur Recovery	2,701	3,144	2,746	—	2,552	2,628	2,498
700 Piping	1,353	1,353	1,353	1,353	1,353	1,353	1,353
800 Utilities	565	934	536	529	535	592	624
900 Waste Disposal	424	424	424	424	424	424	424
Total Direct Costs	21,941	22,472	22,224	18,540	22,002	22,206	22,420
Total Capital Costs	28,699	29,477	29,075	24,208	28,838	29,108	29,452
1977 Annual Costs							
Coal	2,889	2,889	2,889	2,901	2,859	2,889	2,889
Electricity	736	412	787	652	721	734	751
Operating Labor and Materials	1,488	1,488	1,488	1,098	1,488	1,488	1,488
Present Values 1977							
First Year Construction	4,147	4,260	4,201	3,498	4,167	4,206	4,256
Second Year Construction	7,538	7,743	7,637	6,358	7,575	7,646	7,736
Third Year Construction	10,289	10,568	10,424	8,679	10,338	10,435	10,559
Coal	35,442	35,442	35,442	35,589	35,074	35,442	35,442
Electricity	10,354	4,349	9,620	7,498	8,706	8,875	8,988
Operating Labor and Materials	9,679	9,679	9,679	7,142	9,679	9,679	9,679
Product Costs							
Total Project Present Value	77,458	73,492	78,454	70,437	76,974	77,734	78,233
Gas Production, 25 years (Billions of Btu)	48,242	44,794	48,272	49,275	47,183	47,955	47,434
Discounted Average Gas Cost ($ 1 million Btu)	1.61	1.64	1.63	1.49	1.63	1.62	1.65

Gasifier Type: Fluidized Bed
Reactor Mode: Oxygen/Steam Blast

Table F-3

CASE 3 AND VARIANTS
SUMMARIZED INVESTMENT AND OPERATING COSTS FOR VARIANTS
(Thousands of Dollars)

Run Number	1	4	7	10	13	16	19
Run Conditions	NOM	50 PSI	250 PSI	0.5% S	4% S	$e(SO_2)=.6$	$e(SO_2)=.2$
1977 Capital Costs							
100 Coal Preparation	2,228	2,228	2,228	2,228	2,228	2,228	2,228
200 Oxygen Supply	1,774	1,774	1,774	1,782	1,766	1,774	1,774
300 Gasification	7,216	7,216	7,216	7,216	7,216	7,216	7,216
400 Compression	3,350	1,058	4,190	3,417	3,326	3,350	3,332
500 Desulfurization Dehydration	1,622	3,548	1,353	348	2,296	2,359	3,076
600 Sulfur Recovery	1,780	1,875	1,750	—	2,187	1,819	1,828
700 Piping	1,492	1,492	1,492	1,492	1,492	1,492	1,492
800 Utilities	486	514	490	481	490	486	488
900 Wast Disposal	424	424	424	424	424	424	424
Total Direct Costs	20,372	20,129	20,917	17,388	21,425	21,143	21,858
Total Capital Costs	27,106	26,972	27,832	23,026	28,591	28,247	29,314
1977 Annual Costs							
Coal	3,313	3,313	3,313	3,339	3,280	3,313	3,313
Electricity	245	275	353	157	244	263	293
Operating Labor and Materials	1,488	1,488	1,488	1,098	1,488	1,488	1,488
Present Values 1977							
First Year Construction	3,917	3,897	4,022	3,328	4,131	4,082	4,236
Second Year Construction	7,120	7.085	7,310	6,048	7,510	7,420	7,700
Third Year Construction	9,718	9,669	9,978	8,255	10,250	10,127	10,509
Coal	40,644	40,644	40,644	40,963	40,239	40,644	40,644
Electricity	3,440	3,801	4,961	1,649	3,433	3,696	4,062
Operating Labor and Materials	9,679	9,679	9,679	7,142	9,679	9,679	9,679
Product Costs							
Total Project Present Value	74,518	74,838	76,594	67,385	75,242	75,648	76,890
Gas Production, 25 years (Billions of Btu)	48,184	47,953	48,206	49,275	46,920	47,982	47,534
Discounted Average Gas Cost ($ 1 million Btu)	1.55	1.56	1.59	1.37	1.60	1.58	1.62

Gasifier Type: Fluidized Bed
Reactor Type: Air/Steam Blast

Appendix G

COAL GASIFICATION REACTORS

Three types of coal gasification reactors of well known commercial background are available in the U.S. They represent three basically different methods of contacting solids with gases, but all are single-stage reactors. They occupy a special position in coal conversion technology because of their long previous history. However, none of the reactors has been demonstrated commercially in this country. While second generation reactors under development in the U.S. may be more efficient, these also remain undemonstrated.

The gasification chemistry underlying the three systems is fundamentally the same. The coal is partially oxidized by free oxygen and steam, the latter acting as an efficient temperature control agent as well as a source of oxygen. A limited oxidation is sufficient to gasify the solids. If this reaction takes place at high temperature, the resulting combustibles are low in molecular weight, corresponding essentially to CO, CO_2, H_2, H_2O. If, otherwise, the hot gases are allowed to equilibrate partially with cold feed solids, the gases are enriched by the devolatilization of the coal. In part, these options distinguish the gasifiers.

FIXED BED GASIFIER

The fixed bed gasifier is the oldest system and is perhaps the easiest system to operate. It is simply a vessel with provision for feeding solids at the top, for holdup of solids in the middle, and for ash removal at the bottom. The bed of solids is supported on a moving grate. The hearth zone above the grate is the combustion zone, reaching

perhaps 2400°F. Ash liberated in this zone discharges through the grate. Steam and oxygen (or air) supporting the gasification of coal enter at the bottom of the reactor and flow up through the grate and hot reaction zone. The partial oxidation of the carbon here is sufficient to gasify essentially all the carbon, hydrogen, and sulfur, producing primarily CO_2, CO, H_2, and some H_2S in mixture with unconverted water vapor. As the hot gas rises through the cooler layers of descending coal, it volatilizes a substantial hydrocarbon fraction of this "burden." Thus, the gas emerging from the reactor at perhaps 1200°F carries additional components of widely varying molecular weight and condensing temperatures.

The gas is quenched separately in a water wash, producing a certain amount of tar and oils, of the order of 10 percent by weight of feed coal. Some hydrocarbon gases remain after the wash; the heating value of the gas in the oxygen blown case is then in the range of 350-450 Btu/scf.

ENTRAINED BED GASIFIER

The entrained solids gasifier is perhaps the latest of the three reactor types. It is designed to house a reaction zone of high temperature, 3000-3300°F, and a zone of relatively low temperature upstream of a waste heat boiler section. Coal as fine solids is entrained with a steam/oxygen jet and carried into the reaction zone, where the temperature reaches perhaps 3000°F. The coal solids are consumed within a few seconds, and the ash melts and separates as slag. The reaction products are then quenched partially with water spray, to about 2000°F, before passage through a heat exchanger section.

A distinctive feature of the gasifier is that the inventory of solids is small, being in the form of fines suspended in gas.

The flow of solids and steam/oxygen mixture is essentially co-current in character, subjecting all the coal fractions to the same high temperature reaction conditions. The reaction products are free of condensables and contain no hydrocarbons in significant concentration. The walls of the reactor are water cooled to withstand the elevated gas temperature and to recover heat.

FLUIDIZED BED

The fluidized bed technique of contacting gas with solids, now used for many purposes, was first introduced for the gasification of coal. The bed is comprised of relatively well crushed solids — minus 1/8 inch, at least — maintained in a state of "fluid" motion by a rising stream of gas. The motion of the solids is random and sufficiently vigorous to maintain good uniformity of temperature throughout the bed even though local heat effects may be severe. This feature distinguishes the fluid bed reactor from the other types discussed above. It is practical to conduct gasification under conditions in which the temperature is kept below the ash softening point but high enough to provide good reaction rates. This condition allows ash to be removed as "dry," nonagglomerated solids.

The vessel for housing the process is lined with refractory and is fitted with a grate for supporting the bed of solids. A mixture of steam and oxygen (or air) is admitted below the grate, supporting the bed motion and serving the gasification reactions. Fines are lifted out of the turbulent fluid bed with the product gases, which leave the vessel at its top. The fines are returned almost entirely to the reactor, however, after separation from the product gas.

The gas is essentially free of hydrocarbons, since these are not stable chemically at temperatures of the reactor, i.e., 1800 to 2200°F.

TURNDOWN CAPACITY

All the gasifiers have a turndown capacity of 65 to 75 percent; in some cases, this capacity is 100 percent, if some degradation of performance is acceptable. The fixed bed and the fluid bed gasifiers can be shut down and restarted if the inactive interval does not exceed a few hours and if some variation in gas quality is acceptable. The inventory of solids is large enough to maintain combustion temperatures. The entrained solids gasifier on the other hand is cooled by a water-wall, which continues to function in the event of shutdown. It is necessary here to maintain at least a small combustion activity in the reactor to prevent the onset of severe, uneven contractions in the reactor refractory.

SENSITIVITY TO CAKING

The fixed bed and the fluid bed systems are sensitive to coal caking tendency, a condition that leads to clogging of the vessel. A "free-swelling" index of more than 3.5 has been regarded as an indication of unacceptable caking effects. Some vendors claim that provision for breaking up agglomeration can be provided by a rabble arm which rotates slowly through a fixed bed system. No such provision is available for the fluid bed system. Agglomeration tendency can be judged by laboratory tests but should be confirmed by small pilot tests.

The Damm index, used in Germany as a measure of cake strength, can be used as a preliminary measure of coal agglomerative tendency in the fluid bed.

COMPARISON OF GASIFIERS

Characteristics of the commercial gasifiers are summarized in Table G-1, including some features that cannot be discussed within the limitations of this appendix. The operating features of any one gasifier tend to be balanced in net advantage by a different group of features in another.

An important question to be answered is whether the operator is willing to accept a substantial fraction of condensables as part of the product emerging from the fixed bed gasifier. If so, he may produce a richer gas, but will have to perform more processing of both water and the condensable product than otherwise. Furthermore, recovery of heat is only partial.

Table G-1

COMPARISON OF COMMERCIAL GASIFIERS

Feed Coal	Fixed Bed	Entrained Bed	Fluidized Bed	Effects
• Moisture allowable	20%	2%	20%	Drying requirement
• Particle size				
Maximum Minimum	4-6 inches 1/4 inch	[illegible]% minus [illegible] mesh	100% minus 1/4 inch	Grinding requirement
• Air blown mode	Yes	No	Yes	Air blast option
• Caking tendency	Clogs unless stirred	No clogging	Clogs vessel	—
Waste heat recovery	No	Yes	Yes	Recovery of heat
Tar and pitch production	Yes	No	No	Byproduct disposal
Condensables in quench water	Substantial	Negligible	Negligible	Phenol to be removed
Water need	Large	Moderate	Moderate	Water supply
Heating value of gas, Btu/scf	400-450	300	280	O_2 blown
Carbon conversion	0.95	0.95	0.8-0.9	Coal use efficiency
Characteristic temperatures, ^oF	2400/1200	3000/2000	2200/2000	Molecular weight of product gas

A second feature to be considered is safety in operating the plant. In this case, the possibility of oxygen "breakthrough" is a danger. This event arises if there is a momentary disappearance of coal inventory despite a continuation of oxygen feed. The condition is most liable to occur in the entrained solids system, because the inventory can disappear in a matter of seconds if the solids feed is halted. Careful operation of a reliable control system is a necessity. On the plus side of this system is the ability to operate on any coal, a capability which is not possessed by the conventional versions of the other two reactors.

A third feature to be considered is the general flexibility and ease of operation of the fluid bed system, if a non-caking coal is the feedstock. The reactor accepts coal of moderate size, including the fines, at moisture content up to 20 percent. It produces a dry soft ash and a product gas containing no condensables. Its turndown procedure is continuous, and the reactor can be shut down completely for delayed restart without substantial danger to the vessel. The disadvantages of the system are the somewhat low conversion of solid carbon to gas and the sensitivity to caking tendency of some coals.

CHOICE OF THE ENTRAINED SOLIDS AND THE FLUID BED GASIFIERS FOR NAVY APPLICATION

The choice of the entrained solids and the fluid bed reactors for detailed study in this work is based on:

- The small size of the plant
- The absence of condensables in the product gas from these reactors

The complexity of the cleanup system needed to service the output from the fixed bed gasifier is at odds with the low capacity of the system.

A substantial water and product gas cleanup is needed. The processing calls for the scrubbing of sulfur components from the gas and the removal of soluble toxics from the water used to quench the raw gas from the gasifier. To this must be added the disposal or consumption of the tars and oils produced in the gas quench.

As opposed to the above operation, the cooldown of raw gas from either of the two other gasifiers proceeds without condensation of liquids and can take place therefore in a boiler for heat recovery purposes.

In the interest of holding the complexity of the gas and water cleanup to a minimum, particularly for a small plant, the two reactors capable of producing a gas free of condensables were therefore considered to be the logical choices for use.